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Waste Activated Sludge Processing

Los Angeles County Sanitation Districts
Whittier, CA

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16. ABSTRACT

A study was made at pilot scale of a variety of processes for dewatering and stabilization of waste activated sludge from a pure oxygen activated sludge system. Processes evaluated included gravity thickening, dissolved air flotation thickening, basket centrifugation, scroll centrifugation, aerobic digestion, and anaerobic digestion (mesophillic and thermophillic). In addition combinations of processes were evaluated including: scroll centrifugation after anaerobic digestion, basket centrifugation after anaerobic digestion, centrifugation of mixtures of anaerobically digested primary sludge and anaerobically digested waste activated sludge, centrifugation after thermal conditioning and composting after digestion and centrifugal dewatering. Data are presented on all of the processes above and optimum economic combinations are identified.

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August 1980

WASTE ACTIVATED SLUDGE PROCESSING

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FOREWORD

The Environmental Protection Agency was created because of increasing public and government concern about the dangers of pollution to the health and welfare of the American people. Noxious air, foul water, and spoiled land are tragic testimony to the deterioration of our natural environment. The complexity of that environment and the interplay between its components require a concentrated and integrated attack on the problem.

Research and development is that necessary first step in problem solution and it involves defining the problem, measuring its impact, and searching for solutions. The Municipal Environmental Research Laboratory develops new and improved technology and systems for the prevention, treatment, and management of wastewater and solid and hazardous waste pollutant discharges from municipal and community sources, for the preservation and treatment of public drinking water supplies, and for minimizing the adverse economic, social, health, and aesthetic effects of pollution. This publication is one of the products of that research; a most vital communications link between the researcher and the user community.

Federal laws mandating that all POTW plants provide at least secondary treatment will result in significant increases in quantities of waste activated sludge. This report covers a study of a variety of methods of dewatering and stabilizing waste activated sludge.

Francis T. Mayo, Director
Municipal Environmental Research
Laboratory

ABSTRACT

This research program was conducted to determine the most effective means of handling waste activated sludge from the future secondary treatment facilities at the Joint Water Pollution Control Plant in Carson, California. Various methods of thickening, stabilization, conditioning, dewatering, and drying have been evaluated.

Gravity thickening was found to be too sensitive to plant upsets to be a viable thickening option. Dissolved air flotation effectively thickened waste oxygen activated sludge to 3.5% total solids at polymer dosages below 4 lb/ton (2g/kg). Basket and scroll type centrifuges achieved higher thickened sludge solids, but the necessary polymer dosages were higher and the thickened sludge was too viscous to allow mixing during subsequent processes. Disc-nozzle centrifuges were found to be operationally unsuitable for sewage sludge thickening.

Both aerobic and anaerobic digestion were evaluated. Aerobic digestion achieved lower volatile solids destruction than anaerobic digestion and demonstrated no advantages over anaerobic digestion. Thermophilic (120°F or 45°C) anaerobic digestion achieved higher volatile solids destructions than mesophilic (94°F or 34°C) anaerobic digestion, but the increased heating demands more than offset the increase in gas production. No other benefits of thermophilic digestion were found, so mesophilic anaerobic digestion appears to be the most attractive stabilization process.

Vacuum and pressure filtration were evaluated for dewatering the digested waste activated sludge. These processes required extremely high lime and ferric chloride dosages (700 to 800 lb/ton (350 to 400 g/kg) CaO and 250 to 400 lb/ton (125 to 200 g/kg) $FeCl_3$), and polymer conditioning was ineffective for filtration. The filter press produced cakes up to 40% total solids, but the vacuum filter gave wet cakes with poor discharge characteristics.

Scroll centrifugation of anaerobically digested waste activated sludge produced 15% cakes at a 15 lb/ton (7.5 g/kg) polymer dosage. This cake was plastic in nature, but it was conveyable. Basket centrifugation achieved comparable results, but the run times were extremely short. Combining the digested waste activated sludge with digested primary sludge before dewatering may reduce the chemical costs and will result in more manageable cakes. A 70% waste activated sludge - 30% primary ratio is optimum, and separate digestion before combining the sludges seems to be better than digesting the two sludges together.

Thermal conditioning of waste activated sludge greatly improved its dewatering characteristics. Thermally conditioned sludge produced vacuum filter cakes of 31 to 37% total solids and filter press cakes of 34 to 51%. These cakes were solid in texture and easily conveyable. Centrifuge dewatering of thermally conditioned waste activated sludge, however, produced 20 to 22% cakes which were too fluid to be conveyable.

Thermal conditioning produces a sidestream containing concentrated dissolved organics. Soluble COD's in this sidestream averaged about 15,000 mg/l, and the total dissolved solids ran over 12,000 mg/l. Studies concerning the handling of this liquor were not completed for inclusion in this report. Other problems encountered with the thermal conditioning system included odor generation and mechanical failures.

Successful composting of the dewatered digested waste oxygen activated sludge required the recycling of large volumes of dried compost product to adjust the initial moisture content of the cakes. Indirect steam drying was less successful because of the tendency of the sludge to agglomerate into balls which would not dry on the inside.

One of the two most cost effective sludge disposal systems incorporated dissolved air flotation, mesophilic anaerobic digestion, centrifuge dewatering, compost drying, and sale to a fertilizer manufacturer. The other most cost effective system was dissolved air flotation thickening, thermal treatment, filtration dewatering, and landfill disposal.

This report was submitted in fulfillment of Contract No. 14-12-150 by the Sanitation Districts of Los Angeles County under the partial sponsorship of the U. S. Environmental Protection Agency. This report covers a period from June, 1973, to September, 1976. Further studies are being conducted as of this writing.

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Libby Tortorici is now employed by CM Engineering Associates in Vista, California.

SECTION 1

INTRODUCTION

BACKGROUND

In July 1972, the California State Water Resources Control Board (SWRCB) adopted the State Ocean Plan, followed in October 1972 by the passage in Congress of Public Law 92-500, the Federal Water Pollution Control Act Amendments of 1972. PL 92-500 already established secondary treatment as the minimum treatment level for all publicly-owned treatment works discharging to navigable waters of the United States. The State Ocean Plan, while not specifically requiring secondary treatment as a minimum treatment level, advanced such restrictive effluent standards that secondary treatment was the only cost-effective recourse for the Joint Water Pollution Control Plant (JWPCP).

In compliance with certain provisions of the State Ocean Plan, the Los Angeles County Sanitation Districts submitted on January 15, 1973, the "Ocean Plan Technical Report." The Technical Report outlined measures that had to be taken to achieve compliance with the State Ocean Plan. While the Technical Report was primarily intended to address itself to complying with the State Ocean Plan requirements, recognition was also given to the federal requirement that all wastewaters receive secondary treatment. Thus, it was stated in the Technical Report that the Districts would construct a diffused air biological secondary treatment process at the Joint Water Pollution Control Plant (JWPCP), in conjunction with a comprehensive industrial waste source control program, to achieve compliance with the Ocean Plan and federal requirements.

The Districts acknowledged in the Technical Report that other processes such as physical-chemical treatment, mechanical aeration-air activated sludge, and pure oxygen activated sludge might result in a more cost effective system for compliance with standards, but at the time of the Technical Report there was insufficient information on their performance, costs, design criteria and environmental impacts to consider proceeding with them in lieu of the recommended biological treatment process.

Subsequent to the issuance of the Technical Report, the Districts began pilot plant studies at the JWPCP to evaluate the alternative secondary treatment processes and to confirm the suitability of the recommended process. Two relatively small activated sludge pilot plants (a coarse bubble diffused air system and a high purity oxygen system) and a small physical-chemical treatment pilot plant were established for these studies. All of the pilot plant work has been done with the specific intention of obtaining a cost effective full scale system that will achieve an effluent quality in compliance with the Ocean Plan and Federal requirements.

The results of the physical-chemical pilot work were contained in a Districts' summary report entitled Physical-Chemical Treatment Pilot Plant Investigations at the Joint Water Pollution Control Plant. The report concluded that conventional physical chemical treatment was cost prohibitive in comparison to biological activated sludge treatment. This comparison was made neglecting the reclamation of chemical coagulants from sludge, the potential air quality problems associated with the regeneration of carbon, the inability of carbon adsorption to produce an effluent of 30 mg/l BOD_5, and assuming the suitability of the present JWPCP solids processing scheme.

The results of the evaluation of the biological units are contained in the summary report entitled Evaluation of Activated Sludge Pilot Plants at the Joint Water Pollution Control Plant, April 1974. In general, it can be said that the small scale units acknowledged the biological treatability of the JWPCP primary effluent; however, both units proved to be too small in scale to provide meaningful design and operation data for a full scale treatment plant at the JWPCP. To provide this essential information, it was decided that biological pilot units of a significant scale (approximately 0.5 MGD*) would be constructed at the JWPCP. The two 0.5 MGD pilot units consisted of a deep tank (20-25' SWD) air activated sludge system utilizing a mechanical submerged turbine for air diffusion, and a high purity oxygen aeration system. The accompanying clarifiers were rectangular in shape and were of the Districts' conventional clarifier design. It was planned that the evaluation of these units would have provided the vital information required for complete design of a cost-effective full scale biological treatment system. However, in early 1975 the SWRCB embarked upon a greatly accelerated grant construction program, and a major impact was the requirement for completion of plans and specifications for 100 MGD of secondary treatment capacity at the JWPCP by September 1976, with the design of an additional 100 MGD to be completed nine months later. With such an imposing deadline, the data from yet to be completed pilot plant investigations was combined with available information from the literature and the results of a nationwide field investigation by District personnel. The total information developed indicated that the optimum secondary treatment system should be a high purity oxygen activated sludge system utilizing surface aeration equipment and a cryogenic air separation plant for oxygen generation.

While a process selection was made for the aeration system, considerable uncertainty existed concerning the choice of a waste activated sludge processing system. Clearly, the composite evaluation of a biological treatment process for the JWPCP must address the questions associated with the processing of excess biological solids. Handling and disposing of sludge is a major expense in wastewater treatment and the problem becomes particularly acute with the activated sludge process. Historically, waste activated sludge has been a difficult commodity to thicken or dewater because of the large amount of bound water in the cellular mass and the highly concentrated electrostatic charge on the cell wall.

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*This report was prepared with U.S. customary units and then modified to EPA format. Conversions from U.S. to metric units will be found in Appendix B.

PURPOSE AND SCOPE

It is the purpose of this report to set forth the program that has been conducted to determine the optimum method(s) of handling waste activated sludge at the JWPCP, and to report the findings of those studies, which were conducted from June 1973 to September 1976.

A system flow diagram illustrating the assorted unit processes that were investigated in the waste activated sludge processing study is shown in Figure 1. The initial efforts of the study were directed toward thickening of the waste activated sludge. Gravity thickening, centrifugation, and air flotation have been investigated for this purpose. Aerobic digestion, anaerobic digestion, and thermal conditioning were evaluated for their abilities to reduce the solids mass for disposal and to improve the dewaterability of the sludge. Dewatering of digested sludge, thermal conditioned sludge, and thickened waste activated sludge was accomplished by centrifugation and filtration following appropriate chemical conditioning. One possible solution is to add the waste activated sludge to the existing sludge processing system at the JWPCP. The effects on dewatering of combining the primary and waste activated sludge before and after digestion were investigated. Heat drying and air drying/composting of the dewatered sludge has been evaluated.

Certainly, it is to be recognized that there are other unit processes than those shown in Figure 1, such as those involving multiple effect evaporators: sludge/solvent mixtures; pyrolysis and various types of incineration. A majority of these processes are being scrutinized on paper by the Districts and most certainly all by the Los Angeles - Orange County Regional Sludge Management Study. It would seem though, that at the writing of this report either existing local air pollution control standards preclude their use or that their state of development is such that their feasible full scale implementation would be at a date far beyond that required for the initial 200 MGD of secondary treatment at the JWPCP.

STUDY LOCATION

The ideal location for the conduct of these investigations would be at the JWPCP, using the waste activated sludge generated from the 0.5 MGD activated sludge pilot plants. However, when the need for these studies was evident, the biological pilot plants were in the conceptual design stage and far from operational, and the accelerated CSWRCB construction grants program was certainly not anticipated. Moreover, even if the pilot units had been operational it was felt that the amount of sludge to be generated from the pilot plants would not be of a sufficient quantity so as to allow for the examination of sludge processing at a significant enough scale that the results could be the sole criteria for a full scale design. Considering the time problem with the construction of the activated sludge pilot plants and the limitation on the scale of the sludge processing that could be examined with the amount of waste activated sludge that would be generated, it was decided that the Saugus-Newhall Water Reclamation Plant (District 26 Water Renovation Plant) should be used as the site for the initial phase of the waste activated sludge processing studies. With the facilities at this plant, it is possible to conduct thickening, digestion (aerobic and anaerobic), and dewatering studies of waste activated sludge on a large enough scale for the development of significant operational and design parameters.

The Saugus-Newhall WRP is located in Saugus, California. The present plant average flow is 3.2 MGD. The treatment system as shown in the attached Figure 2 involves grit removal, primary sedimentation, activated sludge secondary treatment utilizing a step feed aeration pattern, and chlorination of the secondary effluent. The present solids handling system consists of centrifugal thickening of the waste activated sludge followed by anaerobic digestion (primary and secondary) of the primary and waste activated sludge in two separate sets of digesters. One set of digesters was isolated from daily plant operation and used in the research investigation of the anaerobic digestion of waste activated sludge. The various thickening, conditioning, and dewatering equipment investigated was located in the area labeled "Research Site", while the necessary project staff was housed in the mobile office facility. Preparation of the Research Site included the purchase and installation of appurtenant equipment such as pipes, pumps, and power supply.

Even considering the benefits of scale and digestion facilities existing at the Saugus-Newhall WRP, a legitimate concern arises as to the results obtained from a study using the waste activated sludge at this plant and their direct application for the design of a waste activated sludge handling system at the JWPCP. In addition to the differing influent characteristics of the two plants, the activated sludge system at the Saugus-Newhall WRP is of the conventional Districts' design of step aeration using coarse bubble air diffusion; while the activated sludge plant at the JWPCP will be of a high rate nature, employing pure oxygen gas. How these differences would reflect themselves in the processing of the waste activated sludges is at the least extremely difficult to anticipate. The waste activated sludge processing studies were, therefore, transferred to the JWPCP when the 0.5 MGD biological treatment pilot plants were operational. The work at the JWPCP emphasized those processes which had demonstrated feasibility in the Saugus-Newhall WRP studies.

SECTION 2

CONCLUSIONS

THICKENING

A. Concentrating waste activated sludge can effectively be accomplished by either dissolved air flotation or centrifugation through basket or scroll type centrifuges.

1. Dissolved air flotation will effectively thicken oxygen waste activated sludge to a concentration of 3.5 percent TS with 99+ percent suspended solids recovery. Polymer dosages of 2 to 4 lb/ton are required and solids loadings as high as 4 lb SS/hr-ft^2 can be applied to the flotation cell.

2. Basket centrifugation of oxygen waste activated sludge through a 48" unit will yield cake solids of 5 to 8 percent TS with 95 percent suspended solids recovery. Polymer dosages in excess of 5 lb/ton are required and the average basket run time approximates 13 minutes. Deceleration and knife insertion must be employed due to the dryness of the cake near the basket wall. The effective solids loading has been determined to be 280 lb SS/hr.

3. Scroll centrifugation was not employed to thicken oxygen waste activated sludge at the JWPCP. Small daily sludge quantities and the limited availability of an 18" x 54" scroll centrifuge did not allow for thickening studies to be conducted. Data collected at the Saugus-Newhall WRP and extrapolated to the JWPCP through comparison with the basket centrifuge data indicate that with the addition of 7-11 lb/ton of cationic polymer discharge solids will fall between 7 and 9 percent TS with 95 percent solids capture. The effective solids loading should approximate 600 lb SS/hr through a 32" x 100" unit.

B. Concentrating waste activated sludge via gravity thickening is feasible only when the sludge exhibits good settling characteristics. Bulking or rising sludge conditions adversely affect gravity thickening.

C. Concentration of waste activated sludge via a disc-nozzle type centrifuge requires prescreens and excessive operator attention.

D. Waste activated sludge concentrated to greater than 6% TS is extremely viscous and contains very little free moisture. At concentrations greater than 6% TS, difficulties may arise in pumping the sludge because of its plastic and viscous nature.

E. The operating parameters of the pure oxygen activated sludge system affect sludge thickening characteristics. The most notable parameter was determined to be power input to the system reactor. An input power reduction of approximately 70 percent to the fourth stage reactor significantly enhanced the thickening characteristics of the waste activated sludge.

STABILIZATION

A. Aerobic digestion of waste activated sludge is an alternative to anaerobic digestion which allows substantial reductions in detention period, but the reduction in volatile solids on Saugus-Newhall WRP waste activated sludge approximated only 25% at hydraulic detention periods of 8 to 13 days and volatile solids loadings of 0.070 to 0.105 lb/ft^3-day. Foam problems continually hampered the digestion process posing severe operational problems. Increasing the air rate so as to maintain a residual D.O. in excess of 1 mg/l did not alleviate the foam problem but did increase the population of nitrifying bacteria, accelerating the nitrification process and depressing the digester pH.

B. Anaerobic digestion of thickened oxygen waste activated sludge was successfully accomplished under mesophilic and thermophilic conditions.

1. Under mesophilic operation (93°F), anaerobic digestion destroyed an average of 32 percent of the applied total volatile solids and yielded 14.8 cubic feet of total gas per pound of volatile solids destroyed. The gas consisted of 61 percent methane. An average hydraulic detention period of 22 days and a daily volatile solids loading of 0.085 pounds per cubic foot were maintained. Operation of the digester was stable and no major operational problems were encountered.

2. Under thermophilic operation (120°F) an average of 39 percent of the applied total volatile solids were destroyed. Unit total gas production was measured at 17.0 cubic feet per pound of volatile solids destroyed with a methane content of 60 percent. An average hydraulic detention time of 21 days and a volatile solids loading of 0.074 pounds per cubic foot per day were maintained. No major operational problems were encountered and operation of the digester was stable after the transition from the mesophilic to thermophilic temperature regime.

3. Ammonia nitrogen generation under thermophilic operation may result in concentrations that inhibit biological activity. The average ammonia nitrogen concentration encountered while thermophilically digesting oxygen waste activated sludge approximated 1500 mg/l.

4. For the JWPCP thermophilic digestion of oxygen waste activated sludge cannot be justified when the fuel requirements necessary to sustain thermophilic operation are considered. Based on 100 MGD of secondary treatment capacity, mesophilic digestion of oxygen waste sludge will yield a surplus of approximately $9x10^6$ BTU/day, while thermophilic digestion will require the addition of $45x10^6$ BTU/day.

5. Anaerobic digestion of waste activated sludge requires that the digester biological mass first be acclimated in order to assimilate the waste activated sludge feed. Acclimation can be accomplished by gradually increasing the amount of activated sludge feed to the digester and decreasing the primary sludge feed

to the digester.

C. Thermal conditioning also accomplishes stabilization but the process is primarily intended for conditioning.

DIGESTED SLUDGE CONDITIONING AND DEWATERING

Aerobically Stabilized Sludge, Saugus-Newhall WRP

A. Basket centrifugation through a 48" unit yielded maximum cake solids of 10% TS and 99+% suspended solids capture with the addition of 5 to 18 lb/ton of cationic polymer.

B. Centrifugation through a 20" x 62" scroll centrifuge with the addition of 4 to 26 lb/ton of cationic polymer yielded cake solids of 7% to 10% TS. Suspended solids recoveries varied from 80% to 95% at polymer dosage of 15 to 26 lb/ton.

C. Vacuum filtration of the aerobically digested waste activated sludge with the addition of ferric chloride (0 to 300 lb/ton) and lime (0 to 600 lb/ton) produced cakes of 11% to 15% TS while filter yields varied from 0.5 to 1.0 lb/hr-ft^2. Suspended solids recoveries varied from 87% to 97%.

D. Blending of aerobically digested waste activated sludge and anaerobically digested primary sludge followed by centrifugation via a 20" x 62" scroll centrifuge produced discharge cakes from 11% to 15% TS with cake moisture decreasing as the primary sludge ratio increased. Depending on the percentage of digested primary sludge, polymer requirements varied from 7 to 17 lb/ton to produce effluents with suspended solids concentrations of 1500 mg/l or less.

Anaerobically Stabilized Sludge, Saugus-Newhall WRP

The anaerobic decomposition of primary-secondary treatment sludges was carried out under two modes of operation. These were to combine waste activated sludge with primary sludge followed by digestion and to digest a straight waste activated sludge.

A. Various ratios of combined sludge were digested and subsequently dewatered. These ratios varied from 23% WAS - 77% primary to 70% WAS - 30% primary. As the percentage of waste activated sludge increased, the resulting cake solids via centrifugation contained more moisture and centrate quality deteriorated. Maximum cake solids obtained for any of these combined sludge ratios via centrifugation (basket and scroll type) approximated 13-14% TS and required approximately 15 lb/ton of cationic polymer to produce a centrate containing less than 1500 mg/l of suspended solids. Vacuum filtration of digested combined sludge met with very little success. Maximum obtainable cake solids approximated 8% TS and exhibited extremely poor discharge characteristics. Chemical conditioning prior to vacuum filtration included the addition of cationic polymer (5 lb/ton), ferric chloride (150 lb/ton) and/or lime (400 lb/ton) with resulting filter yields of approximately 0.5 lb/hr/ft^2.

B. Separate sludge digestion followed by blending and centrifugation was also investigated. Blends ranging from 100% WAS - 0% primary to 0% WAS - 100% primary were subjected to centrifugation preceded by polymer conditioning. Discharge cakes decreased linearly from 22% TS to 12% TS as the amount of digested primary sludge decreased and the amount of digested waste activated sludge increased in the blend. Polymer requirements necessary to achieve less than 1500 mg/l of suspended solids in the centrate increased as the ratio of waste activated to primary sludge increased. Polymer requirements for a straight waste activated sludge approximated 35 lb/ton while a straight digested primary sludge required 6 lb/ton.

Mesophilically Digested Oxygen Waste Activated Sludge, JWPCP

A. The vacuum filtration of this particular sludge was unsuccessful. With the addition of 1200 lb/ton of lime and 350 lb/ton of ferric chloride, the maximum obtainable discharge solids approximated 14 percent TS while maximum filter yields approximated 3.4 lb/hr-ft^2. Captured solids had to be manually scraped from the filtering media, necessitating full time operator attention.

B. Successful pressure filtration required the addition of 700 to 900 lb/ton of lime and 240 to 400 lb/ton of ferric chloride. At these dosages, discharge solids varied from 34 to 40 percent total solids, with corresponding filter yields from 0.25 to 0.44 lb/hr-ft^2. Precoating the filter with 10 lb/ton of diatomaceous earth is an optional operation that improves cake discharge characteristics and may reduce maintenance costs. Suspended solids recovery in excess of 99 percent were consistently obtained.

C. Basket centrifugation met with little success. Flow rates of 50 and 35 gpm were applied to a 48" basket with resultant run times of 5 and 8 minutes, respectively. In both situations the effective flow rate approximated 25 gpm. With the addition of 5 to 15 lb/ton of cationic polymer cake solids varied from 5 to 9 percent TS while the resultant centrates contained 2500 to 1200 mg/l of suspended solids. Approximately 15 percent (2.4 ft^3) of the solids retained in the basket had to be skimmed out and were unconveyable. The remaining solids were easily plowed and although plastic in nature were conveyable.

D. Centrifugation through a pilot scale (18" x 54") scroll centrifuge yielded discharge solids of 15 percent TS with the resultant centrate containing 1200 mg/l of suspended solids. Chemical conditioning with 15 lb/ton of cationic polymer was required while flow rates of 10 and 15 gpm were evaluated. Centrifugation of digested waste activated sludge from the Saugus-Newhall WRP through the above unit, under the same operating conditions, yielded cake solids of 10 percent TS with centrate suspended solids approximating 2000 mg/l.

Combination of Mesophilically Digested Primary and Mesophilically Digested Oxygen Waste Activated Sludge

A. Dewatering of digested primary and digested waste activated sludge is improved when the sludges have been digested separately instead of blended prior to digestion. Separate dewatering may also give higher overall solids content than if the sludges are combined prior to dewatering; however, dewatered waste activated sludge by itself is very difficult to handle, and it is desirable to blend in some digested primary prior to dewatering in order to produce a sludge that can be efficiently handled.

B. Vacuum filtration of various blends of digested primary and digested waste activated sludge resulted in cake solids ranging from 9 to 18 percent TS. All of the cakes from the blended sludge mixture required manual scraping of the filtration media, rendering the use of vacuum filters unsuccessful for this application. Chemical addition prior to filtration included 190 to 250 lb/ton of ferric chloride and 700 to 1050 lb/ton of lime.

C. Blended sludge ratios varying from 100 percent primary - 0 percent WAS to 0 percent primary - 100 percent WAS were dewatered via the 18" x 54" scroll centrifuge. The 100 percent digested primary sludge feed required 4 lb/ton of cationic polymer to produce a centrate containing 1500 mg/l or less of suspended solids, and yielded a discharge cake of 25 percent TS. As the ratio of oxygen waste sludge was increased to 35 percent, cake solids decreased linearly to 17 percent TS and polymer requirements increased to 6 lb/ton. From 65 percent primary - 35 percent WAS to a 0 percent primary - 100 percent WAS ratio the resultant cakes continued to decrease in a linear fashion while the polymer requirements increased. At the 0 percent primary - 100 percent WAS ratio approximately 15 lb/ton of polymer was needed while the cake solids approximated 15 percent TS.

D. Pressure filtration of various blended sludge ratios indicated that the addition of digested primary sludge improves the handleability of the digested oxygen waste activated sludge. With the addition of 30 percent digested primary, total filter cake solids averaged 28 percent TS whereas a 100 percent digested WAS dewatered to 22 percent TS. Chemical addition for each of the above sludges approximated 200 lb/ton of ferric chloride and 700 lb/ton of lime.

Thermophilically Digested Oxygen Waste Activated Sludge

A. Thermophilic anaerobic digestion of oxygen waste sludge did not improve sludge dewaterability beyond that obtained with mesophilic digestion.

B. Vacuum filtration yielded cake solids that had to be manually scraped from the cloth media. Chemical conditioning with 1000 lb/ton of lime and 300 lb/ton of ferric chloride resulted in cake solids of 11.5 percent TS, suspended solids capture of 82 percent and a filter yield of 0.84 lb/hr-ft^2.

C. Successful pressure filtration required 800 to 1200 lb/ton of lime and 200 to 300 lb/ton of ferric chloride. At these dosages, cake solids approximated 23 to 31 percent total solids with filter yields approximating 0.25 lb/hr-ft^2.

THERMAL CONDITIONING AND DEWATERING

A. Thermal conditioning of undigested oxygen waste activated sludge under the wet oxidation and heat treatment modes of operation resulted in high degrees of solubilization of the particulate organic material. Under wet oxidation conditioning soluble COD concentrations increased from an average of 1550 mg/l to an average of 14,200 mg/l while the dissolved solids concentration increased from 2200 mg/l to an average of 12,200 mg/l. Heat treatment conditioning resulted in an increase in the soluble COD concentration from 1650 mg/l to 15,300 mg/l, while the dissolved solids concentration increased from an average of 2000 mg/l to 12,500 mg/l.

B. The measured degrees of oxidation were inconsistent with the thermal unit operating conditions. No correlation between the thermal operating parameters (temperature, detention time, and air supply) and total COD reductions was established. The inconsistencies observed indicate that accurate performance predictions for a full scale system will be difficult, if not impossible.

C. The measured reductions in total and fecal coliforms through the thermal conditioning process were erratic. The problems surrounding the coliform kill data remain unanswered at this time.

D. Dewatering of thermally conditioned undigested oxygen waste activated sludge produced vacuum filter cakes from 31 to 37 percent and filter yields from 2.3 to 6.6 lb/hr-ft^2. Pressure filtration yielded discharge solids from 34 to 51 percent TS and filter yields of 0.50 to 1.03 lb/hr-ft^2. Dewaterability was observed to increase with increases in reactor temperature and detention time and no significant differences in dewaterability was observed between wet oxidation and heat treatment conditioning.

E. Dewatering of heat treated sludge through a 20" x 62" scroll centrifuge at the Saugus-Newhall WRP required the addition of approximately 8 lb/ton of cationic polymer for the centrate to contain less than 1500 mg/l of suspended solids with resulting discharge cakes averaging 20% - 22% TS.

F. The addition of raw primary sludge prior to thermal conditioning and dewatering did not enhance dewaterability by vacuum filtration. A blend of 23% primary - 77% waste activated sludge from the Saugus-Newhall WRP was subjected to LPO conditioning and vacuum filtration and produced cakes of 30% to 33% TS with filter yields varying from 2 to 3.5 lb/hr-ft^2. Suspended solids recovery approximated 97%.

G. Anaerobic digestion of oxygen waste sludge prior to thermal conditioning adversely affected sludge dewaterability. Pressure filtration of the anaerobically digested thermally conditioned sludge produced cakes of 30 percent TS with filter yields approximating 0.60 lb/hr-ft^2. This is compared to a filter press cake of 39 percent TS and a filter yield of 0.94 lb/hr-ft^2 for undigested thermally conditioned oxygen waste activated sludge. A fine line exists between the solids content of a dewatered thermally conditioned sludge and the handleability of the resultant cake solids. At a solids concentration of 30 percent TS, the filter press cakes are dry on the outer surfaces but contain a liquid core which is not conveyable. At solids concentrations in excess of 34 percent TS, the filter cake is consistently dry and firm and easily conveyable.

H. Numerous operational problems were encountered with the pilot thermal conditioning unit. These included, scaling of the heat exchange surface, compressor failures, boiler malfunctions and corrosion of the air lines. Odor generations at the JWPCP were less severe than encountered at the Saugus-Newhall WRP because of the use of a wet scrubber and carbon adsorber. However, even with this equipment for vent gas treatment, strong odors were detectable.

SLUDGE DRYING

A. Composting of dewatered digested oxygen waste sludge was successful but because of the high moisture content of the cake solids, the addition of large volumes of compost material was required to bring the initial moisture content of the compost pile to 65 percent or less. The required drying time and final product moisture content approximated that for digested primary sludge.

B. Indirect steam drying of digested dewatered and undigested dewatered oxygen waste activated sludge will require 1300 to 1500 BTU's per pound of water evaporated. Problems were encountered while drying dewatered biological solids because these sludges tended to agglomerate into 2" to 4" balls that would dry on the outside but remained moist and spongy on the inside. The final moisture content after four hours of drying at a jacket temperature of 297°F approximated 65 percent and odors normally associated with thermal conditions were detectable during these drying studies.

SYSTEMS EVALUATION

A. Based on pilot and full scale data and engineering judgment as to the feasibility of certain processes, twenty (20) alternate waste activated sludge handling schemes were established and analyzed for cost effectiveness. For a system to be considered feasible, the overall suspended solids removal had to be in excess of 95 percent and the dewatered discharge cakes had to be of such a consistency to be conveyable and handleable.

It must be realized that all of these systems were selected without significant regard for any future system considerations or ultimate disposal projects that would result from the studies presently being initiated by the Los Angeles County/Orange Metropolitan Area Sludge Management Study.

B. Of these twenty alternatives, the four most cost effective systems are listed below. Cost estimates reflect a consumer price index of 170 and an ENR index of 2400.

1. Flotation-Anaerobic Digestion-Centrifuge-Compost-Fertilizer Mfr. ($95 to $97/ton)

Composting of dewatered digested oxygen waste activated sludge and dewatered blended digested oxygen plus digested primary sludge was successfully accomplished during these studies. If sufficient land area exists for composting and the final product is acceptable to a fertilizer manufacturer, the most cost effective system would include flotation thickening, anaerobic digestion, centrifugation of a 70 percent WAS - 30 percent primary sludge blend, composting and disposal to a fertilizer company. Basket or scroll centrifuges can be employed for the dewatering process. If basket centrifuges are used, the unit cost for this sludge train would be $95 per day ton of solids processed while a unit cost of $97 per dry ton will be incurred if scroll centrifuges are employed.

2. Flotation-Anaerobic Digestion-Centrifuge-Composting-Landfill ($100 - $102/ton).

If composting is practiced but a sufficient market does not exist or the final compost product is not acceptable to a fertilizer manufacturer, this final product would be disposed of at a sanitary landfill. The sludge train would then be flotation thickened, anaerobic digestion, centrifugation of a 70 percent WAS - 30 percent primary sludge, composting and landfill disposal. The unit cost for this system would be $100 and $102 per dry ton of solids processed for basket and scroll centrifugation, respectively.

3. Flotation-Anaerobic Digestion-Dewater-Landfill ($112 - $123/ton)

If composting cannot be accomplished because of land restriction, the most cost effective systems involving digestion would include flotation thickening, anaerobic digestion, pressure filtration or scroll centrifuge dewatering and landfill disposal. The unit cost for the system involving pressure filtration of the digested oxygen waste activated sludge would be $123 per ton of solids processed. The unit cost for the scheme incorporating scroll centrifugation of a 70 percent WAS - 30 percent primary blended digested sludge would be $117 per dry ton. This figure may be reduced to $112 per dry ton processed pending a uling as to whether the 15 percent TS cake to be disposed of is classified as a liquid or a solid. The current practice at the Districts'landfill site is to charge $2.50 per ton for sludges with total solids concentration greater than 25 percent and $3.50 per ton for sludge with solids concentrations less than 25 percent TS.

4. Flotation-Thermal Treatment-Dewater-Landfill ($96 - $97/ton)

The most cost effective sludge handling schemes involving thermal treatment would include flotation thickening, thermal treatment, vacuum or pressure filtration and sanitary landfill disposal. Anaerobic treatment of the liquid side streams associated with thermal treatment would also be incorporated in this particular sludge train. The unit cost associated with these schemes have been estimated at $97 and $96 per ton of solids processed for vacuum and pressure filtration, respectively, and reflect the cost of anaerobic filtration of the liquid side streams.

Composting of dewatered thermally treated waste activated sludge was not considered at this time. The composting characteristics of thermally treated sludge remains for evaluation and there are serious concerns regarding the probable production of odors upon turning of this particular compost material. Additionally, it is not known if the final product would be acceptable to a fertilizer manufacturer if in fact the thermally treated sludge were amenable to composting.

SECTION 3

RECOMMENDATIONS

Based on the results of the studies reported herein, it has been recommended that dissolved air flotation be adopted for thickening of waste activated sludge. Additional work is needed to optimize operating parameters for sludge from the pure oxygen process.

Mesophilic anaerobic digestion of 100% waste activated sludge has been recommended for stabilization. Future work should include repetition of the startup procedure and long-term steady state operation to address potential problems such as foaming and scale formation.

The thermal conditioning studies raised many questions that must be answered before the process could be effectively applied. Further studies should investigate the mitigation of odors, treatment of the high COD side stream, and solutions to operational problems such as corrosion and scaling.

Scroll centrifuges are the most promising of the dewatering devices tested. However, performance of this type of equipment is known to be dependent on the size of the machine, and evaluation of full scale machinery is needed. Entirely new equipment, in particular belt filter presses, have entered the market and deserve thorough evaluation.

Composting plays a major part in current disposal practices at the Joint Water Pollution Control Plant, but composting of waste activated sludge has received only cursory evaluation. Further study of the windrow composting method for processing waste activated sludge, as well as evaluation of new methods such as the static aerated pile, is necessary before a disposal system incorporating composting could be recommended.

The Los Angeles/Orange County Regional Sludge Management Study is still in progress, and further work may be dictated by the results of this study.

SECTION 4

PROCESS RESULTS

THICKENING

As illustrated in Figure 1, three separate unit operations--dissolved air flotation, gravity thickening, and centrifugation--were investigated for thickening waste activated sludge. The waste activated sludge from the Saugus-Newhall WRP was taken from the reaeration tank instead of the return sludge line. The construction of the Saugus-Newhall WRP system would have made obtaining the waste activated sludge from the return sludge line costly and mechanically difficult, and while no evaluation was conducted at that time to determine the relative merits of thickening reaerated sludge as opposed to return sludge, it is believed that the differences are minimal.

The efforts at the JWPCP were focused on thickening waste activated sludge from the 0.5 MGD high purity oxygen activated sludge (UNOX) pilot plant, although some work was conducted using the sludge from the 0.5 MGD mechanical air activated sludge pilot plant. Waste oxygen activated sludge was obtained either from the final stage of the reactor or from the return sludge line. Waste air activated sludge was obtained from the return sludge line.

Gravity Thickening

Gravity thickening is commonly accomplished in a sedimentation tank in which solids separate from the liquid phase by gravity forces and the settled solids are concentrated by the action of gravity and by virtue of the weight of the overlying solids (compaction). Conventional sludge collecting mechanisms with vertical pickets are employed to stir the sludge gently, thereby opening up channels for the release of water and promoting densification.

A 22" diameter by 72" high gravity thickener with vertical pickets, Figure 3, which was operated in a batch manner, was evaluated at the Saugus-Newhall WRP. Sludge bulking is a recurring operational problem at that plant, and a bulking sludge does not favor compaction in gravity settlers. No data were obtained regarding the ability of gravity settlers to thicken waste activated sludge (WAS), but the process was shown to be extremely sensitive to plant upsets such as bulking or rising sludge. Because of process instability, no gravity thickening studies were conducted at the JWPCP.

Dissolved Air Flotation

The flotation process has long been employed in industry especially in mining and refineries for two-phase separation. Generally, the process is applied to systems where there is a large concentration of insoluble or immiscible particles suspended in a bulk liquid. Either the suspended particles are quite small and nearly colloidal or they have a density comparable to that of the bulk liquid. Air bubbles are introduced into the system to combine with the particles resulting in an aggregate with a density sufficiently less than the bulk density to effect flotation and concentration. Flocculation aids such as polyelectrolytes are often used to aid in clarification and concentration. The flotation unit may be either rectangular or circular in design and a dissolved air system may employ either pressurization of the waste stream and/or recycled effluent.

Three dissolved air flotation units were obtained for these studies. Two of these units were rectangular in design and had flotation areas of 14 and 50 ft^2. The third unit had a 6 ft diameter flotation cell.

The two rectangular units were similar (length/width = 2.3) except that the 50 ft^2 unit had more sophisticated controls and a higher head recirculation pump. As shown in Figure 4, the influent WAS solids enter the unit at the bottom via a distribution box where they are blended with a pre-pressurized recycled effluent stream. The recycled stream is pumped to a retention tank that is maintained at 45 to 55 psig in the 14 ft^2 unit and 55 to 70 psig in the 50 ft^2 unit. Air is introduced into the retention tank via an air compressor and the entire contents are continually recycled by a reaeration pump that augments the dissolution of air into the liquid. Following a short retention period, the pressurized air-saturated liquid is discharged to the distribution box through a back-pressure regulator valve and released at atmospheric pressure. The pressurized stream and influent blend in the distribution box with the minute air bubbles adhering to the WAS solids and causing the solids to rise to the surface and be skimmed by the scraper arms.

The first part of this evaluation was conducted at the Saugus-Newhall WRP and consisted of optimizing the 14 ft^2 unit with regard to retention tank pressure as controlled by the recycle rate. With a WAS feedrate of 20 gpm and polymer added in the range of 4 to 16 lb/ton, recycle rates of 16 and 32 gpm (80% and 95% recycle) corresponding to retention tank pressures of 55 and 45 psig respectively, were evaluated. Data for optimization of the unit is shown in Figures 5 and 6. As shown in Figure 5, varying the recycle rate from 16 to 32 gpm had very little effect on the concentration of the float solids but suspended solids recovery was greatly affected by the recycle rate as shown in Figure 6. At a recycle rate of 16 gpm, suspended solids (SS) recovery in excess of 99% were consistently obtained whereas at a recycle rate of 32 gpm SS recovery ranged from 92% to 95%.

It can be argued that a poorer solids capture obtained at the higher recycle rate (32 gpm) was attributed to excessive turbulence through the unit due to an increased hydraulic loading but data presented by Hayes[1] provides a further explanation.

According to Hayes, a linear relationship exists between the percent saturation of air in water and the detention time in the pressurized retention tank. As the retention tank pressure increases, the recycle flow decreases and consequently the detention time in the pressurized retention tank increases. As the detention time increases, the percent saturation increased by virtue of the reaeration pump and consequently a greater number of minute air bubbles are released when the pressurized recycle flow is released to atmospheric pressure. With this increase in the number of minute air bubbles, the attachment of air to solids in the distribution box is enhanced because as Ettelt[2] has pointed out, the smaller bubbles have less liquid to displace from the surface of the solids to which they must attach and, therefore, they attach more readily than larger bubbles. Additionally, because their terminal velocities are less than those of larger bubbles, the detention time is also increased which appreciably enhances the opportunity for contact with the solids thus allowing for more solids to float to the surface of the unit and be removed by the scraper mechanism.

With the unit operating at the lower recycle rate of 16 gpm, a retention tank pressure of 55 psig and 0.6 lb/hr of air added to the retention tank, the second part of the evaluation was conducted. The effects of solid and hydraulic loadings on float solids concentration and solids capture were investigated over a wide range of polymer dosages. These results are shown in Figures 7 and 8. It should be noted that with this particular unit the maximum amount of air added to the retention tank was limited to 0.6 lb/hr because of limitations on the air injection system. As seen in Figure 7, the float solids concentration did not vary as the solids loading increased from 2.3 to 6.2 lb/hr-ft^2 corresponding tc hydraulic loadings (excluding recycle) of 0.75 to 1.5 gpm/ft^2, respectively. As the solids loading increased to 9.3 lb/hr-ft^2 (2.6 gpm/ft^2 excluding recycle), a pronounced drop in float solids concentration is observed at polymer dosages of 7-10 lb/ton, but at polymer dosages of 11-13 lb/ton the variation in float solids concentration is insignificant. With respect to solids capture, increases in solids loading and hydraulic loadings over the range investigated had very little effect on suspended solids removal.

As shown in Figure 8, the fact that a total hydraulic loading of 3.8 gpm/ft^2 including a recycle of 16 gpm still produced suspended solids removal in excess of 99% adds support to the theory presented earlier. For example, at a total hydraulic loading of 3.8 gpm/ft^2 (Figure 8), solids recovery in excess of 99% were obtained because the lower recycle rate yielded a longer retention tank detention time and a degree of air saturation.

The efforts extended towards evaluating the 50 ft^2 rectangular dissolved air flotation unit were limited because the unit was being utilized strictly as an operations tool and as such only a limited number of operating parameters were investigated. In fact, only one waste activated sludge feedrate was applied to the unit. At a feedrate of 63 gpm, corresponding to a hydraulic loading of 1.2 gpm/ft^2 and a solids loading of 5 lb/hr/ft^2, polymer was added in the range of 4 to 7 lb/ton and the A/S ratio was maintained at 0.018. At a total hydraulic loading of 3 gpm/ft^2 (including recycle) the unit consistently removed 99% of the influent suspended solids (Figure 8) and the resultant float solids obtained are shown in Figure 7.

During the flotation studies, the Saugus-Newhall WRP sludge volume index (SVI) averaged 360 ml/g with a range of 215 ml/g to 615 ml/g. These high SVI's are indicative of poor sludge thickening qualities.

In September 1975, the 14 ft^2 rectangular flotation unit was relocated to the JWPCP research site. The flotation studies conducted at the Saugus-Newhall WRP utilized waste activated sludge taken directly from an aeration tank. The relative merits of thickening mixed liquor as opposed to return waste sludge could not be evaluated at that time, so the initial phase of the JWPCP studies included a series of tests on both pure oxygen mixed liquor from the final stage of the UNOX reactor and waste sludge from the return sludge line.

In addition to comparing the flotation characteristics of mixed liquor and return waste sludge, these initial studies investigated the effects of storing return waste sludge for up to 12 hours prior to flotation. The relatively low wasting rate from the 0.5 MGD pure oxygen pilot plant and occasional operational problems which caused variations in the wasting rate necessitated storing sufficient quantities of excess sludge to ensure a constant loading rate to the flotation cell. It, therefore, became necessary to determine if this practice of storing sludge prior to flotation was detrimental to its flotation characteristics.

The results from this evaluation are summarized in Figures 9 and 10 which reflect the data collected at feed rates of 5 and 8 gpm, respectively, to the flotation cell. For these runs, the flotation unit was operated at a retention tank pressure of 55 psig and a recycle rate of 25 gpm with the addition of 0 to 11 lb/ton of cationic polymer. The results indicate that the oxygen sludge thickening characteristics did not vary significantly with sludge origin of storage up to 12 hours. More important, these results confirm that the differences in flotation performance between a return sludge feed and a mixed liquor or reaerated sludge would have been minimal at the Saugus-Newhall WRP.

Evaluation of the rectangular dissolved air flotation unit on oxygen waste activated sludge continued on a regular basis for approximately eight months after these initial studies. During the duration of the studies, the performance of the flotation unit fluctuated considerably and performance criteria were established to help characterize the flotation characteristics of the sludge as operational parameters varied within the pure oxygen system. Based on float solids concentration, SS recovery and chemical requirements, flotation performance was categorized as either good, marginal, or poor. Those runs which met the criteria of 3.0 to 4.0 percent float solids, and 99 percent SS recovery with 0 to 4 lb/ton of polymer were characterized as good, while marginal or poor performance indicates that one or more of the performance criteria were not met.

The initial operation of the air flotation unit was conducted during the start-up period of the 0.5 MGD pure oxygen activated sludge system. Because of the start-up problems with the pilot plant and the many operating variables inherent to any secondary biological system, it was extremely difficult to correlate any single operating parameter with subsequent flotation of the waste activated sludge. On a very general basis, the waste activated sludge exhibited good floating characteristics when the unstirred SVI was below 85, the MCRT was less than six days, the sludge blanket level in the final clarifier was less than five feet, and the total system solids approximated 3600 pounds or less. Periods of marginal or poor performance were normally encountered when the above parameters exceeded the specified limits. The results from the operation of the air flotation unit during this period can best be described as sporadic, and in general inferior to what has been reported in the literature by pure oxygen system manufacturers. Figures 11, 12, and 13 represent flotation performance during periods of good, marginal, and poor operation, and include the summary of the pure oxygen system and the flotation unit operation parameters.

On April 30, 1976, the speed of the surface aerator in the fourth stage of the oxygen system was reduced from 68 to 45 rpm, resulting in an input power reduction of approximately 70 percent. By May 6, 1976, the flotation unit performance was visually observed to improve and on May 7, 1976, a program was set up to ascertain if the oxygen waste sludge would consistently concentrate to within the limits of the established performance criteria. For 19 days in a six-week period, the flotation unit was closely monitored. The results from this evaluation are presented in Table I which includes a summary of the pure oxygen system and flotation unit operating conditions. The established performance criteria were consistently met during these evaluations and a review of the oxygen system operating parameters indicates that the power density change in the fourth stage reactor enhanced the flotation characteristics of the waste sludge. The oxygen system parameters encountered during these studies ranged over the full spectrum of conditions encountered prior to the power reduction; yet the flotation performance was consistently categorized as good. The power change reduced shearing forces within the fourth stage reactor resulting in a waste sludge more amenable to flotation thickening and less sensitive to system operating variations.

A comparison between the data presented in Table I and Figure 7 indicates that the oxygen WAS from the JWPCP will flotation thicken better than the WAS from the Saugus-Newhall WRP. At polymer dosages of 4-6 lb/ton, the Saugus-Newhall sludge yielded float solids at about 3.5 percent TS. The oxygen WAS gave an average float solids of 4.1 percent TS and a minimum of 3.3 percent TS at polymer dosages less than 3.2 lb/ton.

The six-foot diameter circular flotation cell, as furnished by the manufacturer, was divided into three separate compartments or flotation cells by vertical steel baffles that were slotted at the bottom and terminate approximately 6" below the liquid surface at the top. Each cell contained a one horsepower pump, a back-pressure regulator valve, and an air aspirator system which was the sole source of air for solids flotation.

This was the only pilot circular flotation unit available for rental and it was certainly unique and would not be considered typical of a standard circular flotation unit for use on waste activated sludge.

To make the unit suitable for waste activated sludge, changes were made by Districts personnel and the unit operated under these changes will be hereinafter referred to as the "modified" flotation unit. Since the modified unit more closely resembles a standard circular flotation cell, only data collected under the modified mode of operation will be presented and discussed. The modified system, as shown in Figure 14, employed coupling pumps 1 and 3 in a series arrangement along with the addition of a retention tank downstream of the pumps. The WAS stream entered the suction of pump 3 where it combined with the underflow of Cell 3. Air was then added at the discharge side of Pump 3 by means of an air compressor which replaced the venturi aspirator system. The air-liquids-solids mixture then passed through Pump 1 and was discharged to a retention tank operating at approximately 60 psig. Following a short retention period, the pressurized mixture was discharged into Cell 1 at atmospheric pressure through a modified inlet works which included a perforated clay distribution box.

With all the modifications made, the unit was still limited by the surface sludge scraper system. It was visually ascertained that at the solids loadings in excess of approximately 3.6 lb/hr-ft^2 (22 gpm) the scraper mechanism was not able to recover solids at the same rate they were captured. To correct this would have required extensive alterations and, as such, the unit was operated with this limitation. In addition, in view of the fact that Cell 3 was utilized only as the draw off of clear underflow, it can be concluded that only two-thirds of the unit was utilized for flotation. Therefore, the effective surface area used was 18.67 ft^2 instead of 28 ft^2, and the actual maximum solids loading was 5.4 lb/hr-ft^2 instead of 3.6 lb/hr-ft^2. With an air to solids (A/S) ratio averaging 0.023 and cationic polymer added in the range of 4 to 16 lb/ton for conditioning, composite samples of various runs were taken and the results are shown in Figures 15 to 16.

It should be noted that although there was a pressurized retention tank to aid in the dissolution of air, this system was quite inferior to the air injection system on the rectangular units evaluated. Specifically, the retention tank pressure could not be easily controlled and the recycle rate could only be estimated to be between 100% and 200% of the feedrate. In addition, the lack of a mixing device in the retention tank limited the degree of saturation obtainable. As presented by Eckenfelder[3], the use of mixing in a pressurized retention tank can produce 90% of saturation whereas 50% of saturation is usually obtained in an unmixed pressurized retention tank.

The plot of float solids versus polymer dosage, Figure 10, shows that a maximum float solids concentration of 3.5% TS was obtained on this modified unit with the addition of 13 lb/ton of polymer. SS recoveries in excess of 99% were obtained at all the polymer dosages as shown in Figure 16, but at the lower range of 4 lb/ton, float solids of only 2.3% TS were obtained.

The circular dissolved air flotation unit was not evaluated on oxygen WAS.

Centrifugation

The centrifuge is not new to wastewater treatment; sanitary engineering literature since the beginning of the century is sprinkled with reports of sludge centrifugation. A perforated basket-type was used in Germany to dewater raw primary sludge as long ago as 1902; and in Milwaukee a centrifuge was evaluated in 1920 but operating results were disappointing.[4] Only in recent years have centrifuges come into fairly common use, however, and factors which have contributed to the increase in the number of centrifuge installations include alteration of centrifuge design to make the machines more suited to use with the types of solids encountered in waste treatment and the availability of synthetic organic polyelectrolytes for sludge conditioning.

Basically, centrifuges separate solids from the liquid through sedimentation augmented by centrifugal force. Sludge is fed into the rotating bowl at a constant feedrate where it separates into a dense cake containing the solids and a dilute centrate stream containing fine, low-density solids. Three different classes of centrifuges were evaluated for sludge thickening: 1) basket centrifuge, 2) horizontal scroll, and 3) disc-nozzle. The results obtained are given below:

Basket Centrifuge--

Three 48" diameter, imperforated bowl, basket centrifuges were examined for WAS thickening. Each of these units rotated at approximately 1380 rpm which is equivalent to 1300 gravities at the bowl wall. Except for minor variations in the chemical injection system and the feed inlet works, the units were identical baring differences in the drive mechanisms.

The basket centrifuge is a solid bowl which rotates along a vertical axis and operates in a batch manner. A schematic of a basket centrifuge is shown in Figure 17. The feed material is introduced at the bottom of the unit and is accelerated radially outward to the wall of the basket through centrifugal force. Cake continually builds up within the basket until the quality of the centrate, which overflows a weir at the top of the unit, begins to deteriorate. At that point, feed to the unit is stopped and a skimmer enters the bowl to remove its contents. The total solids concentration of the cake increases in an outward radial direction, and as a result cake solids concentrations near the basket wall can be of such a magnitude as to prevent their being removed by the skimmer. The field operations indicated that when the WAS is thickened to a composite basket cake concentration of less than 6% TS full depth skimming is possible while the basket is revolving at full speed (1380 rpm). At this speed, the basket acts as a centrifugal pump and the skimmings are discharged through a hose. Upon completion of the skimming sequence, which takes one minute, the feed sequence is again initiated. When the composite cake is greater than 6% TS, those solids near the basket wall are too thick to be skimmed and, as a result, to remove this material the machine must be decelerated and the remaining cake plowed out. The need to plow out the cake is a major detraction of this unit, when compared to either the air flotation or horizontal scroll continuous discharge centrifuge.

In total, the three machines produce competitive results and, as such, distinctions were not made as to manufacturer in the discussion of data. The machines were operated at hydraulic feedrates of 50 gpm, 65 gpm, and 70 gpm, with cationic polymer added in the range of 0 to 27 lb/ton. As can be seen from Figure 18, when the cake solids concentration is plotted as a function of polymer dosage, a well defined single curve is obtained. The thickened solids ranged from approximately 4% TS with no chemical conditioning to a maximum of approximately 8% TS with 25 lb/ton of cationic polyelectrolyte. As noted previously, in general observation, when the composite thickened solids concentration was less than 6% TS, the material could be completely removed from the machine via the skimmer nozzle; however, in excess of 6% TS the solids near the basket wall had to be knifed. The majority of the data points shown in Figure 18 were duplicated several times, and particularly for the single point shown for 65 gpm. On a great number of occasions the machines were run at 65 gpm for extended periods of time to supplement the existing sludge thickening mechanism at the Saugus-Newhall WRP. The results from these runs were very consistent and, as such, were averaged and shown as the single data point. The corresponding solids recoveries and centrate SS concentrations are shown in Figure 19. As can be seen, rather than define a single function, the data appears to divide hydraulically between those runs at 70 gpm and those at 50 and 65 gpm. For 50 gpm, the solids recovery ranged from 80% (centrate = 1200 mg/l SS) with no polymer addition, to in excess of 95% (centrate = 300 mg/l SS) at polymer dosages greater than 15 lb/ton. The duplicated runs at 65 gpm yielded results that were equivalent to the 50 gpm data. For 70 gpm, the results at polymer dosages in excess of 15 lb/ton were, for all practical purposes, equivalent to those for 50 gpm. However, as the polymer dosage was decreased the effluent quality for the 70 gpm condition decreased at a faster rate than for the 50 gpm condition. It is difficult to clude if the deteriorating effluent quality is a result of either hydraulic or solids loading rate limitations, or a combination of both.

One of the 48" basket centrifuges was moved to the JWPCP to evaluate various polymers in conjunction with the second stage sludge dewatering station. Scheduling difficulties, operational problems with the test centrifuge, and a limited WAS supply greatly restricted the thickening studies on the oxygen WAS, but such data as were collected are summarized in Figures 20 and 21.

The unit was operated at a constant bowl speed of 1380 rpm, corresponding to an acceleration force of 1300g's and was loaded at a constant hydraulic rate of 50 gpm. Polymer addition in the range of 0 to 5 lb/ton yielded composite cake solids of 6 to 8 percent TS while SS recovery varied from 77 to 95%. Because the composite cake was greater than 6% TS, full depth skimming could not be employed. The volume of sludge skimmed out is proportional to the variation in solids content as the skimmer moves into the captured solids. When the skimmer reaches a point where the solids concentration is in excess of approximately 7% TS, it will no longer advance into the cake and the machine has to be decelerated and the ploy inserted to remove the remaining solids. The deceleration and plow insertion sequence in conjunction with average run times of 13 minutes yielded an effective flow rate of 38 gpm and suspended solids loadings of 280 lb/hr. These data were collected when the oxygen WAS exhibited good flocculation and flotation characteristics and, therfore, reflect optimum performance.

The Saugus-Newhall WAS required 10-12 lb/ton of polymer, whereas the oxygen WAS required only 5 lb/ton to obtain 95 percent SS recovery. The data indicate that these differences in performance are not due to differences in sludges as much as the SS recovery really does not describe the performance of a centrifuge as well as centrate SS does. Both sludges gave centrate SS concentrations of about 700 mg/l at the 5 lb/ton polymer dosage. With the oxygen WAS (1.47 percent) this centrate corresponded to a 95 percent recovery; but with the Saugus-Newhall WAS (0.55 percent), the recovery was only 87 percent. A SS recovery of 95 percent with the Saugus-Newhall WAS requires a centrate with only 300 mg/l SS.

Horizontal Scroll Centrifuge--

Two horizontal scroll, concurrent flow, tapered bowl centrifuges were evaluated as WAS thickening devices. Both machines were manufactured by the same company; however, they were of different sizes (32" bowl diameter x 100" bowl length vs. 20" x 62") and had different hydraulic capacities. A schematic of the basic characteristics of each machine is shown in Figure 22.

The scrolls of each machine rotate along a horizontal axis and operate in a continuous manner. Sludge is fed to the unit through a stationary tube along the centerline of the inner screw which accelerates the sludge and minimizes turbulence. The sludge passes through ports in the inner conveyor shaft and is distributed to the periphery of the bowl. Solids settled through the liquid pool in the separating chamber are compacted by centrifugal force against the wall of the bowl and are conveyed by the outer screw conveyor to the opposite end of the inlet works. Separated liquid (centrate) is discharged continuously over an adjustable weir at the inlet end.

The machines were evaluated at different time periods during the study. For all data presented utilizing the 32" x 100" centrifuge, the bowl speed was maintained at 1280 rpm (750g's). The relative scroll speed was held constant at 16.5 rpm's and the pool depth was maintained at maximum. When utilizing the 20" x 62" centrifuge, speeds of 2070 gpm (1200g's) and 19 rpm were maintained for the bowl and relative scroll, respectively, and the pool depth was maintained at maximum. Preliminary testing with each unit governed the relative scroll speed, while the desire to obtain maximum solids recovery set the pool depth. The machines were operated at mid and low pool depths to ascertain the effect on centrate quality and cake solids, but the change was negligible.

The results from the operation of the 32" x 100" unit are presented in Figures 23, 24, and 25. The hudraulic flow rate ranged from 70 to 90 gpm while the resulting solids loading rates varied from 185 to 280 lb SS/hr. As shown in Figure 23, these variations in loading rates did not affect the cake solids concentration, nor did the polymer dosage have a significant effect on the concentrations obtained. The cake solids concentration varied from approximately 7% TS with no polymer addition to 8% TS with the addition of polymer at a rate of 20 lb/ton. However, for the corresponding solids recovery and centrate concentration, as shown in Figures 24 and 25, the hydraulic and solids loading rates, as well as the polymer dosage had a significant effect. In general, at cationic polymer dosages greater than 15 lb/ton there was no significant difference in centrate quality or solids recovery for the hydraulic and solids loading rates encountered. However, at dosages less than 15 lb/ton it can be seen that as the solids loading rate increased the centrate quality and solids recovery decreased.

In the operation of the 20" x 62" unit, the flow rate was varied from 40 to 60 gpm, while the corresponding solids loading rate ranged from 72 to 108 lb SS/hr. As shown by the results in Figures 26 and 27, the same general operational response was observed in this unit as compared to the larger 32" x 100" unit. While no data was obtained at polymer dosages of less than 5 lb/ton, the results indicated that a cake of 7.5% TS could be obtained at 7 lb/ton of polymer and could be increased to approximately 9% TS with a polymer dosage of 25 lb/ton. These results were slightly better than that achieved with the larger machine, although quite obviously at much reduced hydraulic and solids loading rates. As illustrated in Figure 27, the suspended solids recoveries obtained with the smaller machine approximated those achieved with the larger unit and demonstrated the same response in regard to increased solids loading rate. Both machines required cationic polymer dosages in excess of 10 lb/ton to achieve adequate solids recoveries (95%), with the corresponding cake solids in the range of 7% to 9% TS. As such, the choice of which machine to implement on a full scale application would result from the flow and quantity of solids to be handled in combination with an economic analysis of the thickening system.

Scroll centrifuge thickening of the oxygen WAS was not evaluated. Assuming that the centrate SS vs. polymer dosage curves for the two WAS's will be similar for the scroll centrifuges as they were for the basket centrifuges, it is possible to extrapolate the Saugus-Newhall data and estimate the results of scroll centrifuge thickening at the JWPCP. In order to obtain 95% SS recovery on the oxygen WAS, a centrate with 600 mg/l SS is required. The 32" x 100" scroll centrifuge required 7 lb/ton of polymer and the 20" x 62" machine required 11 lb/ton to meet this criterion with the Saugus-Newhall WRP WAS. These same dosages would be expected at the JWPCP. The resultant cake solids would be at least as high as the 7-9% TS cakes produced in the Saugus-Newhall WRP studies.

Disc-Nozzle Centrifuge--

The disc centrifuge has a perforated bowl which rotates along a vertical axis at approximately 6000 rpm and operates in a continuous manner. A schematic of the disc centrifuge is shown in Figure 23. The feed material is introduced at the top of the unit and flows through a set of some 50 conical discs which are utilized for stratification of the waste stream to be clarified. The discs are fitted quite closely together and centrifugal force is applied to the relatively thin film of liquor and solids between the discs. This force throws the denser solid material to the wall of the centrifuge bowl where it is subjected to additional centrifugal force and concentrated before it is discharged through nozzles located on the periphery. The clear liquid continually flows over a weir at the top of the bowl and exits via the centrate line. The bowl is equipped with 12 nozzle openings, but various numbers and sizes of discharge nozzles can be utilized depending on the feed liquor and the desired results. The number and size of discharge nozzles used directly influences the sludge concentration for any given feed condition.

Historically, prescreening of the feed material has been of necessity, because of the machine nozzle size (0.07 in. to 0.08 in.). As such, screens with openings of .030" and .027" were installed upstream of the centrifuge but failed to successfully remove small sand particles that eventually clogged the nozzles and continuously interrupted operation of the unit. It was the intent to study the machine's thickening capabilities over a full range of WAS flows with and without polymer conditioning. However, because of the prescreening problems, the evaluation of the machines was greatly curtailed. Perhaps with the implementation or development of adequate prescreening mechanisms the machine could realize its potential; however, for this investigation it was decided that the disc-nozzle mechanism was simply not competitive with the previous systems evaluated.

Process Selection

The thickening performance data are presented in Table II. Table II reflects data collected at the JWPCP on oxygen WAS for dissolved air flotation and basket centrifugation. The scroll centrifuge data are projected to the oxygen WAS based on the scroll centrifuge results on the Saugus-Newhall WAS and the basket centrifuge data from both research locations. No usable data could be obtained for gravity thickening or disc-nozzle centrifugation.

Centrifugation will consistently produce cakes of 6 to 8 percent TS. WAS solids at that concentration are very plastic and viscous in nature, and it is the opinion of the authors that conventional gas recirculation mixers will not provide adequate mixing for the digestion process if the digesters are fed 6% WAS.

Dissolved air flotation is the most attractive alternative. It will provide sludge thickened to 3.5 percent TS at polymer dosages less than 4 lb/ton. The SS recoveries will be greater than 99 percent. The high SS recovery is especially beneficial for the operation of the secondary treatment system since it will allow more accurate control of the system solids. Also, those solids which would escape from the thickening system will be the most difficult to handle, and returning them to the main treatment stream would only contribute to operational problems. Dissolved air flotation will minimize those problems.

Thickening systems will be further discussed in the cost analysis section of this report.

STABILIZATION

The satisfactory disposal of the concentrated organic solids removed from sewage in the primary sedimentation tanks and excess biological solids from the activated sludge process frequently requires that the solids first be stabilized. The original objective of stabilization was to reduce the objectionable qualities of the sludge such as putrescibility and odors. In practice, some of the side benefits of stabilization may become the primary objectives. The basic benefits of stabilization that were of interest in these studies were volatile suspended solids destruction, energy production, and improved dewaterability of the sludge.

The most common and widely used method of sludge stabilization is anaerobic digestion where the decomposition of organic and/or inorganic matter is performed by microorganisms in the absence of molecular oxygen.

Other unit processes commonly employed for the stabilization of sewage sludges include aerobic digestion and oxidation ponds. Thermal treatment involves heating the sludge for short periods of time at elevated temperatures and pressures. Since this process is primarily intended for conditioning it will be discussed in a separate section.

Aerobic Digestion

Aerobic digestion may be defined as the destruction of degradable organic sludges by aerobic, biological mechanisms and has essentially evolved from the extended aeration version of the activated sludge process. The process may be used for either primary sludge, excess biological sludge, or mixtures of the two. Generally, aerobic digestion is most applicable to excess biological sludges because in the absence of an external substrate, microorganisms enter the endogenous phase of the life cycle, resulting in a net decrease in the degradable portion of the microbial or sludge mass.

As an integral part of the biological sludge treatment studies, aerobic digestion of thickened waste activated sludge was investigated at the Saugus-Newhall WRP. A 13,000 gallon, coarse bubble, diffused air system was installed at the research site in November 1974 and served as an aerobic digester for approximately six months. Initially, it was the intent of this evaluation to investigate aerobic digestion of waste activated sludge over a full range of volatile solids loadings, detention periods, and air loadings; but disappointing results prematurely ended the study.

The operating parameters that were investigated are presented in Table III. Detention times of 8 and 12.7 days were investigated while volatile solids loading rates varied from 0.070 to 0.105 lb VSS/ft^3-day with the air input maintained at 0.043 or 0.060 cfm/ft^3.

During the first month of operation, (November 1964) severe foaming in the digester hampered its operation and foam spillage occurred daily, causing a portion of the digester solids to be washed out of the system. It was impossible to calculate the amount of solids leaving the system via the foam. Hence, a true assessment of the volatile solids destruction efficiency was not possible during this period. No direct measures were taken to eliminate the foaming problem but with the addition of approximately 1500 ml per day of defoaming agents the problem was alleviated and the foam confined to the digester.

Tables IV through IX represent the operational and performance data collected during each month of the study and, as seen in Table IV, the residual dissolved oxygen (DO) maintained in the digester during November was only 0.18 mg/l. Relatively little design or operational data are presented in the literature for aerobic digestion but minimum DO concentrations of 1.0 to 2.0 mg/l are normally recommended.

In December 1974, the air loading rate was increased to 0.060 cfm/ft^2, resulting in an average residual DO concentration of 1.05 mg/l. As presented in Table V, at a hydraulic detention time of 8 days and a volatile solids loading of 0.090 pounds per cubic foot per day, an average of 26.8% of the applied volatile solids were degraded. During this period a notable decrease in pH was observed and it dropped as low as 4.9 before two pounds of NaOH were added to suppress the pH decline.

The most plausible explanation for the decline in pH is that excessive nitrification was occurring in the digester. As nitrification becomes more complete, the acidity increases because of a greater number of hydrogen ions going into solution. The increase in nitrification was believed to be related to the higher air rate being maintained to the aerobic digester. When the air rate was increased to 105 cfm, the digester residual DO increased to approximately 1.0 mg/l as opposed to a residual DO of .2 mg/l or less during the month of November. This substantial increase in DO, in conjunction with the other favorable conditions such as higher NH_3 concentrations in the feed, resulted in an inordinate growth or bloom of the nitrifying bacteria and an increased rate of nitrification.

During the next three months; January, February, and March, 1975; the digester was operated at a hydraulic detention time of 8 days, an air loading rate of 0.060 cfm/ft^3 and a volatile solids loading varying from 0.089 to 0.105 lb/ft^3-day. The data collected during this period are summarized in Tables VI, VII, and VIII.

Volatile solids destruction declined slightly during this three-month period to an average of 23%. The residual DO concentration and oxygen uptake rates approximated each other during the months of December and January but during the first of February the DO uptake began to steadily decrease while the digester residual DO began to steadily increase. During this time microscopic examinations revealed that the biological population in the digester was also declining.

By February 20, 1975, the oxygen uptake rate had declined to 14.7 mg/l/hr and the number of rotifers and ciliates (stalked and free swimming) had declined drastically. It should be noted that during this period the plant aeration tank solids had declined significantly and the foam problem had become particularly acute. In fact, the MLSS concentration in Aeration Tank No. 1 had fallen to below 0.3% SS. On February 21, 1975, the step aeration flow pattern was changed. The MLSS concentration in Aeration Tank No. 1 immediately began to increase while the aeration foam problem was alleviated. These changes immediately affected the aerobic digester. By February 25, 1975, the residual DO decreased to 0.2 mg/l, the oxygen uptake rate increased to 45.5 mg/l/hr, rotifers and ciliates again became predominant and the foam in the aerobic digester was alleviated.

During the month of March (Table VIII) the digester operating parameters approximated those of the previous three months and the volatile solids destruction leveled off at an efficiency of 22%. The DO uptake rate increased slightly and the residual DO concentration decreased to 0.20 mg/l while the recorded rate of nitrification decreased from the rates recorded previously. The reason(s) for the decrease in nitrification and residual DO are not apparent but are thought to be related to the step changes made in the aeration system at the end of February 1975.

The digester operating parameters were changed during the month of April 1975, and the data recorded during this period is presented in Table IX. The digester was operated at a hydraulic detention time of 12.7 days, a volatile solids loading of 0.070 lbs VSS/ft^3-day and an air loading rate of 0.060 cfm /ft^3. Under these operating conditions, volatile solids destruction increased slightly to a value of 26.4%. The residual DO increased to 3.5 mg/l while the DO uptake decreased to 37.5 mg/l/hr, indicating that the system could have been operated at a lower air loading rate than the 0.060 cfm/ft^3.

Aerobic digestion studies were not conducted at the JWPCP.

Anaerobic Digestion

In modern practice, anaerobic digestion is usually accomplished in heated reactors maintained within the mesophilic temperature range (90° to 100° F). The elevated temperature has been found to speed up the digestion process and to improve the process stability. Anaerobic digestion can also be conducted in the thermophilic temperature range (120 to 135°F). The organisms involved in thermophilic digestion are not the same as are involved in mesophilic digestion, so the process results may be different.

Both mesophilic and thermophilic anaerobic digestion were evaluated in these studies.

Mesophilic Digestion--

In May of 1973, an unsuccessful attempt was made at the Saugus-Newhall WRP to anaerobically digest straight waste activated sludge. The failure was attributed to not allowing the bacteria time to acclimate. Waste activated sludge was pumped to the digester while completely halting primary sludge pumpings and, as a result, the digester failed and anaerobic digestion of straight waste activated sludge was thought to be impractical. In August 1973, the Research Section at the Saugus-Newhall WRP began monitoring a 125,000 gallon, gas-mixed, heated primary digester which was receiving a combined sludge of approximately 20% waste activated and 80% primary sludge. This digester was isolated for the biological sludge treatment research studies and the waste activated to primary sludge ratio was gradually increased until the digester was fed 100% waste activated sludge.

By April 1974, the digester was successfully digesting a 73% - 27% mixture of waste activated and primary sludge but operational problems which developed at the plant led to a digester failure in July 1974. As a result, the digester had to be reseeded and the ratio of waste activated to primary sludge had to again be gradually increased.

Table X summarizes the anaerobic digestion operating parameters evaluated at the Saugus-Newhall WRP. Included in Table X is a summary of data collected on combined sludge anaerobic digestion at the Valencia WRP. Figure 29 is a plot of volatile solids destruction versus the percentage of waste activated and primary sludge. Volatile solids destruction in excess of 50% were consistently obtained for all of the combined sludge ratios. Digestion of 100% waste activated sludge produced volatile solids destructions of 45-50%.

Unfortunately, valid gas data were not collected in the Saugus-Newhall studies because of the plant's practice of continually hauling stored sludge from the 125,000 gallon secondary digester and an inadequate gas metering system. The gas lines from all four of the digesters at the Saugus-Newhall WRP are interconnected and it was virtually impossible to collect isolated gas data from any of the digesters. In addition, excessive hauling from the secondary digester which often resulted in a blown seal and gas leakage through the seal. Even after an additional gas meter was installed in January 1975, the blowing of seals continually hampered the collection of gas data.

In August 1975, the 13,000 gallon diffused air aerobic digester was transported to the JWPCP and converted to a 12,000 gallon anaerobic digester. The unit served to digest thickened waste activated sludge and was operated in the mesophilic and thermophilic temperature ranges.

In mid-September 1975, the pilot digester was seeded with 5,000 gallons of digested waste activated sludge from the Saugus-Newhall WRP and immediately began receiving air flotation thickened waste activated sludge from the JWPCP pilot plants. Due to various start-up and operational problems, the digester did not reach steady state conditions until the latter part of October 1975. Tables XI and XII summarize the operational parameters maintained and the performance achieved during a 61-day period from November 1975 to January 1976. During this steady state period, the digester received an average of 551 gpd of thickened oxygen and air waste activated sludge. It was the intent of this study to digest oxygen sludge only, but fluctuations in the wasting rate from the oxygen system necessitated that waste sludge from the mechanical air system be added so as to ensure a uniform loading rate to the digester. An exact ratio of oxygen to air sludge added to the digester is difficult to determine and the best estimate is that 20-30 percent of digester feed contained waste sludge from the air system. The volatility of the two waste sludges approximate each other and it is not felt that the addition of air sludge had any significant effects on the project objectives.

At an average hydraulic detention period of 22 days and a volatile solids loading of 0.085 pounds per cubic foot per day, a volatile solids destruction efficiency of approximately 32 percent was recorded in the digester. Total gas production averaged 14.8 cubic feet per pound of volatile solids destroyed and consisted of approximately 61 percent methane. Operation of the digester during this period was stable and no major operational problems were encountered other than normal start-up difficulties.

Subsequent to the collection of data for this report, an inadequacy in the volatile acids analysis at the JWPCP which caused low results was discovered. The average volatile acids concentration reported was less than 10 mg/l for mesophilic digestion.

In addition to the routine analysis reported in Tables XI and XII, one set of samples were taken and analyzed for heavy metal concentrations. These samples consisted of the feed oxygen waste activated sludge and the digester effluent. The results from these analyses are presented in Table XIII. It is important to note that the feed sample consisted of a single grab sample and while the digester effluent sample was also a grab, it actually reflects the accumulation of approximately 20 days worth of storage in the digester. It is the author's opinion that the influent and effluent total metals concentration should be essentially equivalent. That there are some substantial differentials is thought to be reflective of the lack of statistical significance in a single sample. Even considering the sampling procedure it is significant to note that the soluble metals concentrations remained virtually unchanged through the digestion process.

Thermophilic Digestion--

On January 5, 1976, the temperature of the pilot digester began to be increased from 94°F to a targeted thermophilic temperature of 120°F. By January 13, 1976, the digester temperature had reached 103°F with no appreciable changes in any of the performance or operating parameters. In an attempt to retard a digester upset, the digester feed rate was lowered on January 13 to reduce the solids loading rate and to increase the hydraulic detention time from 19 to 29 days. On January 15 the temperature had reached 108°F and except for a slight decrease in methane quality, the performance and operating parameters remained constant until January 22 when the digester temperature was recorded at 113°F. This caused an immediate rise in volatile acids, a further decrease in methane quality and mild foaming conditions. To avert an ultimate failure, digester feed was halted on January 23, resulting in a sharp drop in gas production. The digester was closely monitored with respect to pH, volatile acids, alkalinity and methane gas quality, and the temperature was again increased, to 120°F, on January 26. On the following day, approximately 10 percent of the digester volume was displaced with thermophilically digested sludge from the Hyperion Treatment Plant to establish a healthy thermophilic bacteria population in the pilot digester.

On January 28 the digester began to again receive oxygen waste activated sludge with continued close monitoring of the vital operating parameters. Figure 30 shows the digester response (ph, volatile acids, and methane gas quality) to temperature increases during the transition period from mesophilic to thermophilic operation. Due to the analytical problem mentioned earlier, the volatile acids data are probably inaccurate, but since all analyses were conducted by the same method, changes in the results are significant.

Steady state conditions were achieved by the end of February 1976, and Tables XIV and XV summarize the operational and performance parameters obtained through June 1976. Volatile solids destruction averaged 39.4 percent while the unit gas production was measured at 17.0 cubic feet per pound of volatile solids destroyed. An average hydraulic detention time of 21 days and a volatile solids loading of 0.074 pounds per cubic foot per day were maintained during this four month period. No accurate volatile acids data are available for thermophilic digestion, but the available data and a characteristic odor indicate that thermophilic digestion will result in higher volatile acids than mesophilic digestion. The reported volatile acids concentration averaged 90 mg/l, but due to the previously mentioned laboratory problem, the actual volatile acids concentration is unknown. A characteristic volatile acids odor was detected during the thermophilic study, but not during mesophilic digestion, which confirms the qualitatively higher volatile acids resulting from thermophilic digestion. Again, it was the intent of this program to solely digest oxygen waste activated sludge but due to the problems mentioned previously, the digester feed sludge contained approximately 20 - 30 percent of waste sludge from the 0.5 MGD mechanical aeration air system.

Of interest is the fact that all of the anaerobic digestion studies conducted to date on waste activated sludge, at both the Saugus-Newhall WRP and the JWPCP, indicate that these waste sludges are not destroyed or converted as readily as raw sludge solids in anaerobic digestion. The difference in volatile solids destruction between the two sludge types is most likely related to the relative amount of degradable solids contained in each of the respective sludges and the effects of ammonia toxicity as outlined by McCarty and McKinney.[5]

Ammonia is usually formed in anaerobic treatment from the degradation of wastes containing proteins or urea. Inhibitory concentrations may be approached in industrial wastes containing high concentrations of these materials or in some highly concentrated municipal waste sludges.

Ammonia may be present during treatment in the form of the ammonium ion (NH_4+) or as dissolved ammonia gas (NH_3). These two forms are in equilibrium with each other, the relative concentration of each depending upon the pH or hydrogen ion concentration.

According to McCarty,[6] the following ammonia nitrogen (sum total of the ammonium ion plus ammonia gas) concentrations which may have an adverse effect on anaerobic digestion are listed below.

NH_3N- Concentration (mg/l)	Effect on Anaerobic Treatment
15 - 200	Beneficial
200 - 1000	No Adverse Effect
1500 - 3000	Inhibitory at Higher pH Values
Above - 3000	Toxic

If the concentration is between 1,500 and 3,000 mg/l, and the pH is greater than 7.4 to 7.6, the ammonia gas concentration can become inhibitory. A review of the thermophilic digestion data collected at the JWPCP (Table XV) indicates the digester operated within this region of inhibited biological activity. Unfortunately, nitrogen analyses were not conducted during previous mesophilic digestion studies but it is the author's opinion that higher than normal ammonia concentrations would have been encountered.

If the assumption is made that excess biological solids are not as amenable to anaerobic treatment as raw sludge solids solely because of the buildup of ammonia nitrogen, it would then seem that two solutions are available to alleviate the problem. These are: to dilute the biological sludge feed prior to digestion or to blend concentrate biological solids with primary sludge solids prior to digestion.

In April 1976, a laboratory scale study was initiated to study the effects of biological sludge feed concentrations on subsequent ammonia nitrogen generation in the anaerobic treatment process. At the time of this writing, the laboratory scale digesters were approaching steady state conditions and the results from these studies will be reported at a later time.

In regard to total volatile solids destruction, a review of the digestion data collected at the Saugus-Newhall WRP indicates that approximately 40 percent of the applied volatile solids are destroyed by anaerobic digestion whereas mesophilic digestion of excess activated at the JWPCP yielded 32 percent volatile solids destruction and 39% for thermophilic digestion. The variation in measured efficiency might be related to the fact that the Saugus-Newhall WRP treats primarily a domestic waste while a combination of industrial and domestic waste is treated at the JWPCP. The industrial fraction of the waste treated may be less suited to anaerobic digestion than domestic wastes.

One set of samples consisting of digester feed sludge and digester effluent were collected during the thermophilic studies for heavy metal analysis. These results are presented in Table XVI and indicate a higher degree of solubilization than under mesophilic conditions. It should again be noted that although both grab samples were taken at the same time, the effluent sample reflects 20 days of detention in the digester and may be regarded as a composite while the feed sample was collected over an eight-hour period.

Process Selection

When comparing aerobic digestion and anaerobic digestion, it becomes evident that from a solids reduction standpoint there are no advantages to aerobically digesting waste activated sludge. Additionally, aerobic digestion requires an energy input whereas anaerobic digestion is self-sustaining and produces a usable source of energy. Furthermore, as will be discussed in the following section, the dewaterability of waste activated sludge was not enhanced by aerobic digestion when compared to anaerobic digestion.

Thermophilic anaerobic digestion appears attractive because of a 23 percent increase in volatile solids destruction and a 15 percent increase in unit gas production (ft^3/pound destroyed) when compared to mesophilic digestion, but it cannot be justified when fuel requirements for heating are considered.

Figures 31 and 32 summarize the theoretical heating requirements for mesophilic and thermophilic digestion. These calculations are based on a secondary treatment capacity of 100 MGD and incorporate the average digestion performance results achieved with the JWPCP research digester. Assuming a total system thermal efficiency of 50 percent mesophilic digestion will yield a surplus of 9×10^6 BTU/day, while thermophilic digestion will require an addition of 45×10^6 BTU/day. Additionally, the dewatering properties of thermophilic digested sludge were not observed to increase over those of mesophilically digested oxygen sludge. A complete cost analysis for the various stabilization options will be presented in the system analysis section of this report.

DIGESTED SLUDGE CONDITIONING AND DEWATERING

Subsequent to stabilization, sludges can be conditioned and dewatered so so that their moisture content is considerably reduced. As a result of dewatering, the transportation costs to final disposal are reduced; or if incineration or mechanical drying is practiced, reductions in fuel requirements may be realized. The dewatering system must also achieve an effluent low enough in suspended solids that it can be either returned to the head end of the treatment plant without causing any adverse effects on the operation of the plant or it can be combined with the plant effluent without appreciably affecting the effluent quality.

Two basic mechanisms are employed to dewater sludges. In filtration, a matrix is established which allows water to pass but retains the solids. The driving force can be provided by pressure pumps on the feed sludge or vacuum pumps on the filtrate side. In centrifugation, sedimentation is enhanced by centrifugal force. Chemical conditioning with polymer and/or inorganic chemicals (lime, alum, and ferric chloride) can improve the performance of dewatering equipment.

The research digesters provided aerobically and anaerobically digested WAS and anaerobically digested blends of primary sludge and WAS. In addition to these sludges, blends of anaerobically digested WAS and anaerobically digested primary sludge were available for the entire study, so each machine was tested on a limited selection of these sludges.

Filtration

Pressure Filter (Filter Press)--

Pressure filtration is a batch operation and consists of vertical steel plates (trays) which are held rigidly in a frame and pressed together. A schematic of the plate and frame pressure filter is shown in Figure 33. The sludge is fed into the press and passes through feed holes in the trays along the length of the press. As filtration proceeds, the liquid passes through the fiber of the cloth media and the solids are retained. Sludge feeding is stopped when the cavities between the trays are completely filled. Drainage ports are provided at the bottom of each press chamber, and the filtrate is collected and discharged to a common drain. The dewatering step is complete when the filtrate flow is near zero. The plates are then disengaged and the filter cake is discharged.

Pressure filtration studies at the Saugus-Newhall WRP were confined to thermally conditioned sludge and will be discussed in another section.

The manufacturer of the pilot filter press evaluated at the JWPCP provided a specific filtration resistance meter, a modified Buchner funnel, for preliminary testing of the sludge and its reaction to conditioning chemicals. The specific filtration resistance (R) determinations on mesophilically digested WAS are summarized in Tables XVII and XVIII.

Four series of "R" test were run. The first utilizing a 3.8 to 1 ratio of lime (CAO) to ferric chloride, a ratio that the filter press manufacturer had frequently found to be cost-effective. The following three series of runs were conducted using either ferric chloride, lime, or cationic polymers for conditioning. The equipment manufacturer suggested that an "R" reading of 2×10^{12} cm/g or less is required for effective dewatering on a full scale basis.

As seen in Tables XVII and XVIII, the addition of lime, ferric chloride, or polymer alone did not reduce the R value to the desired range. The addition of 16 percent (320 lb/ton) ferric chloride and 61 percent (1220 lb/ton) lime did reduce the specific filtration resistance reading to below 2×10^{12} cm/g. Although ferric chloride and lime are the standard conditioning agents used in conjunction with vacuum and pressure filtration, an evaluation of polymer conditioning was considered appropriate at least on a laboratory scale level.

The pressure filter manufacturers indicated that polymers "break down" at the high pressures encountered in pressure filtration and as seen in Table XVII, the R readings did increase significantly as the pressure was increased from 100 psig to 225 psig. The increase in the R value was minimal as the pressure increased from 45 psig to 100 psig, indicating that the polymer is strong enough to withstand pressures at least up to 100 psig.

The 30 ft^2 pilot pressure filter was operated at lime dosages (CaO) of 30 percent (600 lb/ton) to 50 percent (1000 lb/ton) and ferric chloride dosages of 8 percent (160 lb/ton) to 20 percent (400 lb/ton) while dewatering mesophilically digested oxygen WAS. The unit was operated at a pressure of 225 psig for filtration periods of 2 to 3 hours. The results of these studies, all of which were run at a 30 mm cake thickness, are summarized in Table XIX. The average feed solids concentration during these studies approximated 3.0 percent TS. The cake solids data shown in Table XIX ranged up to 40 percent at the higher chemical dosages and three-hour run times. It was found that discharge cakes containing less than 30 percent total solids would not discharge completely. Cakes below this point, although firm next to the filter, had a fluid core which allowed the cake to split down the middle when the filter press was open. As a result, the fluid portion would discharge while the firm portion remained attached to the filter media, requiring manual removal and cleaning.

In order to meet or exceed 30 percent total solids in the filter press cake, this particular sludge required lime dosages of 35 percent (700 lb/ton) to 40 percent (800 lb/ton) CaO and ferric chloride dosages of 12 (250 lb/ton) to 20% (400 lb/ton). A precoat application of approx. 10% (200 lb/ton) diatomaceous earth preceded the filtration sequence. Precoating is an optional operation that improves the discharge characteristics and reduces maintenance costs. At these dosages, discharge solids varied from 34 to 40 percent total solids with corresponding filter yields ranging from 0.25 to 0.44 lb/hr-ft^2. For all of these runs, the suspended solids capture exceeded 99 percent.

One filtration run was made with the addition of 20 lb/ton of cationic polymer but as was the case with polymer addition to vacuum filters, its use in this application was unsuccessful and was not further evaluated.

A number of filtration experiments were also conducted on digested primary sludge and mixtures of digested primary and digested waste activated sludge. The results from these evaluations are presented in Table XX. The three runs on digested primary sludge again indicate that the filterability of this sludge is poorer than had previously been recorded.[7] The improvement in handleability of digested waste activated sludge by blending with digested primary sludge prior to filtration was again verified in these tests. A combined mixture of 70 percent WAS - 30 percent primary (Table XX) dewatered to 28 percent total solids with the addition of 720 lb/ton CaO and 200 lb/ton of ferric chloride. At these same chemical dosages, straight digested waste activated sludge (Table XIX) dewatered to 22 percent total solids.

Prior to dewatering of the thermophilically digested oxygen WAS, R measurements were made to indicate the sludge response to conditioning agents. Table XXI summarizes the R measurements with and without preconditioning with lime and ferric chloride. When comparing this data with the R measurements made on mesophilically digested waste activated sludge (Table XVII), it becomes evident that no advantages in dewaterability are realized by digesting in the thermophilic temperature range. Included in Table XXI is the result of conditioning the digested sludge with a body feed of 100 percent (1 ton/ton) diatomaceous earth. This measurement was made at the request of the filter manufacturer based on their success at other installations. The resulting R reading approximated 100 x 10^{12} cm/g and was well above the recommended maximum value of 2 x 10^{12} cm/g. Further investigation into the use of diatomaceous earth was not considered justified because of the high R value and the resultant increase in total mass of solids to be disposed of in subsequent dewatering processes.

The results of filter pressing the thermophilic sludge on the pilot (30 ft^2) filter are presented in Table XXII. The data are presented in chronological order because of observed changes in sludge filterability with time. It should be noted that degradation in filterability of mesophilically digested sludge was not observed, and the change in filterability of the thermophilic sludge may have been attributed to not allowing sufficient time for the digester to stabilize after it had reached its targeted temperature of 120^{0}F. The first six runs presented in Table XXII were conducted prior to two hydraulic detention times and as such, the digester may still have contained a significant quantity of mesophilic sludge. Runs 7 through 12 were conducted after the digester had gone through two detention periods. As seen in Table XXII, the filterability increasingly degraded for runs 7 through 10 as evidenced in the filter yield and cake solids. Although the filter performed satisfactorily at the higher lime dosages used in runs 11 and 12, these dosages were much higher than were required for the mesophilic digested oxygen waste sludge. Additionally, these higher filter yields were attributed to operating at a 2-1/4 hour filtration time as opposed to 3 hour runs for the remaining filter tests. Suspended solids removal for all of these filtration experiments were consistently in excess of 99 percent.

The manufacturer of the 30 ft^2 pilot filter press used for these studies is the only company manufacturing presses that operate at 225 psi. Numerous companies manufacture presses with operating pressures of 100 to 125 psi and investigations to determine the effects of pressure on filtration performance were considered appropriate. These pressure evaluations were conducted on a 0.33 ft^2 prototype filter press that was manufactured and supplied by the same company that supplied the 30 ft^2 unit. The feed sludge for these studies consisted of thermophilically digested waste oxygen activated sludge that had been chemically conditioned with 62 percent lime (1240 lb/ton) and 16 percent ferric chloride (320 lb/ton). As seen in Figure 34, insignificant variations in filter performance are observed when operated at pressures of 225 or 125 psi. The filtrate volume with time was essentially equal for the two operating pressures and the total amount of solids retained were within 5 percent for both operating conditions. Although the cake total solids were higher for the run operated at 125 psi, the discrepancy is most likely due to sampling. The lack of benefit from the higher pressure is attributed to the compressibility of the biological sludge solids. The increased pressure causes an increase in the water's velocity through the cake, but at the same time it compresses the cake and reduces the size of the interstices, and no net increase in total flow is realized.

Vacuum Filter--

Vacuum filtration is a continuous process and consists of a rotating drum which continuously passes through a trough containing the feed sludge. A schematic of the rotary drum belt-type vacuum filter is present in Figure 35. The cylindrical drum which is covered with a cloth media is submerged approximately 20% to 40% in the trough. Radial partitions divide the drum into compartments, each alternately subjected to a vacuum. As a vacuum of 20 to 25 inches of mercury is applied, a sludge mat forms on the filtration media while the filtrate or effluent is discharged. As a point on the filter drum rotates out of the trough, the vacuum is decreased and the captured solids are subsequently removed via the discharge roller.

At the Saugus-Newhall WRP, vacuum filtration studies using the 3' x 1' rotary drum filter were conducted on both aerobically and anaerobically digested sludges. Only one operating condition of the aerobic digester was evaluated. During the period from December 1974 through March 1975, the aerobic digester was operated at a hydraulic detention time of eight days, on average solids loading of 0.095 lb VSS/ft^3 day, and an air loading rate of 0.060 cfm/ft^3.

Chemical conditioning with ferric chloride in the range of 0 to 300 lb/ton and lime in the range of 0 to 600 lb/ton was the only method of condition employed prior to vacuum filtration of the aerobically digested waste activated sludge. The results from this evaluation are presented in Figures 36 through 39.

At a drum cycle time of approximately 2-1/2 minutes, an applied vacuum of 22" Hg and utilization of a tightly woven nylon cloth, cake solids, and filter yields increased with increasing additions of ferric chloride (Figure 36). With no chemical addition, cake solids approximated 11.5% TS with a filter yield of 0.5 lb/hr-ft^2. Cake solids approximated 15% TS and the filter yield increased to 1.0 lb/hr-ft^2 with the addition of 300 lbs/ton of ferric chloride. Solids recovery decreased with increasing amounts of ferric chloride. As seen in Figure 37, solids recoveries of 97% were recorded with no chemical addition while a recovery of 87% was obtained with the addition of 300 lb/ton of ferric chloride. Increases in drum cycle time yielded increased cake solids while filter yields and solids recoveries decreased.

Preconditioning with 0 to 600 lb/ton of lime prior to vacuum filtration was also investigated. Applied vacuums of 21.5 to 22" Hg and drum cycle times from 2 to 7 minutes were employed. As seen in Figure 38, as cycle time increased cake solids increased and filter yields decreased at each of the applied lime dosages. Cake solids varied with chemical dosage and cycle time but little difference in cake solids was observed with chemical dosages from 100 to 600 lb/ton and cycle times between 2.25 to 5.75 minutes. At a cycle time of 2.25 minutes and no chemical addition, cake solids of 11.7% TS and filter yields of 0.48 lbs/hr-ft^2 were observed. The addition of lime from 100 to 600 lbs/ton increased cake solids to 12.5 to 14% TS and increased filter yields to a maximum of 0.55 lbs/hr-ft^2. Increasing the cycle time to 5.75 minutes increased cake solids to 14.5% TS while decreasing the filter yield to 0.3 lb/hr-ft^2. The data collected on solids recovery was scattered with respect to cycle time and lime dosages. At a cycle time of 2.25 to 2.75 minutes, recovery varied from 97.5% with 100 lb/ton of lime to 87% with the addition of 200 to 600 lb/ton of lime. It should be noted that during all these vacuum filtration experiments the solids retained on the filter media exhibited extremely poor discharge characteristics. The depth of solids buildup was small and the cloth had to be manually scraped in order for the solids to be removed from the filter.

The vacuum filter was evaluated on two digested blends of WAS and primary sludge. These blends resulted from the anaerobic digester at the Saugus-Newhall WRP being converted incrementally from primary sludge feed to WAS feed, so the two sludges were blended before digestion.

The 43 percent WAS - 57 percent primary blend was evaluated using two different nylon cloths and various cycle times. Chemical addition ranged from no chemicals to 5 lb/ton of cationic polymer, 125 lb/ton of ferric chloride and 200 lb/ton of lime. The results from this evaluation are presented in Figures 40 and 41. Cake solids varied from 6 to 8.5% TS over the range of parameters investigated and, as expected, cake solids increased slightly and the filter yields (lb/hr-ft^2) decreased with increasing drum cycle times. The filter yields were extremely poor (0.5 to 0.15 lb/hr-ft^2) while solids recovery was consistently greater than 93% and the filtrate contained less than 1500 mg/l of suspended solids. The filter cloths used were tightly woven, which may account for the high moisture content in the discharged cake because of the inclusion of fine solids on the cloth. Coarser cloths would have produced slightly drier cakes but fine solids would have passed through and increased the amount of solids recycled with the filtrate stream. It should be noted that none of the operating conditions produced readily dischargeable cakes. In fact, during each of these runs the cloths had to be manually scraped in order to remove the captured solids.

A similar vacuum filtration evaluation was conducted on the 70 percent WAS - 30 percent primary blend. As shown in Table XXIII, the chemical dosages tested were higher than with the 43 - 57 blend (up to 400 lb/ton of lime and 150 lb/ton of ferric chloride), but the process results were comparable. The filter yields were less than 0.6 lb/hr-ft^2, the cakes were 8.7 percent TS or less, and the cake discharge was poor, requiring the media to be scraped.

The vacuum filter used in the JWPCP studies was similar in construction and the same size, 3' x 1', as the unit tested at the Saugus-Newhall WRP. The mesophilically digested oxygen waste activated sludge exhibited extremely poor vacuum filter dewatering characteristics. The results obtained are summarized in Tables XXIVA and XXIVB. The vacuum filter was operated at cycle times varying from 2 to 6 minutes with an applied vacuum of approximately 23 inches Hg. The maximum obtainable discharge solids approximated 14 percent TS with the addition of 1200 lb/ton of lime and 350 lb/ton of ferric chloride. Filter yields peaked at approximately 3.4 lb/hr-ft^2. Chemical dosages less than those listed above yielded lower discharge solids and reductions in filter yields. The addition of 100 lb/ton of alum and 10 lb/ton of cationic polymer preceded a number of filter runs but poor results renders their use unsuccessful for this application.

In an effort to improve the handling characteristics of digested oxygen waste activated sludge, various dewatering tests were conducted on combinations of digested primary sludge and mesophilically digested oxygen sludge. The digested primary sludge was generated at the JWPCP and three runs were made on this sludge for the sake of background information. The data generated are summarized in Tables XXVA and XXVB. Chemical addition prior to filtration included 190 to 250 lb/ton of ferric chloride and 700 to 1050 lb/ton of lime. Cycle times were varied from 2 to 6 minutes while applied vacuums of 23 inches Hg were maintained.

Resultant cake solids on the digested primary sludge feed approximated 25 percent TS while filter yields varied from 2.5 to 3.7 lb/hr-ft^2 over the range of parameters investigated. The three runs at 100 percent digested primary sludge indicate that the filterability of the digested primary sludge is poorer than in 1973.[7] This decline in filterability has been attributed to an unexplained shift toward smaller particles in the particle size distribution of the incoming sewage and, consequently, the raw and digested primary sludges. The smaller particles have more surface area per unit mass and therefore require higher chemical dosages for coagulation. The digested primary sludge, however, did still dewater better than the digested waste activated sludge.

Combined digested sludge feeds resulted in discharge solids ranging from 9 to 18 percent TS while filter yields varied from 0.80 to 4.5 lbs/hr-ft^2. It should be noted that for all runs made on the combined sludges, the captured solids exhibited extremely poor discharge characteristics and had to be manually scraped from the filter cloth. In those runs classified as "poor", considerable cleaning of the media was also required. The digested primary sludge exhibited "fair" discharge characteristics indicative of self-discharging cakes but considerable cleaning of the media was still required. Runs listed as "good" are indicative of self-discharging cakes with media cleaning not required.

One set of vacuum filtration tests were conducted on thermophilically digested oxygen waste activated sludge, and poor results rendered this dewatering technique unsuccessful. The thermophilic sludge was conditioned with 50 percent lime (1000 lb/ton) and 15 percent ferric chloride (300 lb/ton) prior to filtration. The filter was operated under a vacuum of 18 inches Hg and a cycle time of 4:45. The resultant cake solids were measured at 11.5 percent TS and had to be manually scraped from the belt. The filter yield was calculated at 0.84 lb/hr-ft^2 while 82 percent of the applied suspended solids were captured. A comparison of this data with that presented for mesophilic sludge (Table XXIV) indicates that thermophilic digestion offers no dewatering advantages and further vacuum filtration studies were not considered justified.

Centrifugation

Only basket and scroll centrifuges were evaluated for digested sludge dewatering. These machines were described in the "Thickening" section of this report. Primary sludbe and WAS were combined in various ratios both before and after digestion, so, for clarity, the following notation will be used in this section:

1. Sludges which were blended prior to digestion will be referred to as (% WAS): (% Primary) digested blend.

2. When digested primary sludge was combined with digested WAS or a digested blend, the product will be referred to as (% WAS): (% Primary) combined digested sludges.

Horizontal Scroll Centrifuge--

Three scroll centrifuges were evaluated in the dewatering studies at the Saugus-Newhall WRP. Two of these described as "tapered bowl", scroll centrifuges, were different size models provided by the same manufacturer and measured 32" (bowl diameter) x 100" (bowl length) and 20" x 62". The third machine was provided by a competing manufacturer and measured 14" x 48". The 20" x 62" scroll centrifuge was the only scroll centrifuge available for dewatering the aerobically digested WAS.

From December 1974 through March 1975, the aerobic digester was operated at a hydraulic loading of eight days, an average volatile solids loading of

0.095 lb VSS/ft^3-day and an air loading rate of 0.060 cfm/ft^3. Polymer addition was the only conditioning method employed to aid in centrifugal dewatering. Cationic polymer was added in the range of 4 to 26 lb/ton, while various feed rates (20 to 40 gpm), relative scroll speeds (19 and 24 rpm) and the maximum and middle liquid pool depths were evaluated. For all test runs, the unit was operated at a bowl speed of 2070 rpm's (1200 g's) and the results are presented in Figures 42 and 43. Cake solids varied from 7.5 to 10% TS and were insignificantly affected by the various centrifuge operating parameters. Suspended solids recoveries were more significantly affected by variations in feed rate, relative scroll speed and liquid pool depth. With the exception of the runs made at 20 gpm, which may have received an improperly prepared mixture of chemicals, polymer dosages had to be increased with increases in feed rate and relative scroll speed to produce centrates with suspended solids concentrations of 1500 mg/l or less. A comparison of the data collected at feedrate of 30 gpm and the different pool depths illustrates the effect of pool depth on centrate quality. At corresponding polymer dosages, the runs made at the maximum pool depth consistently recovered more solids than those runs made at the middle pool depth.

During April 1975, the aerobic digester was operated at a hydraulic detention time of 12.7 days, a volatile solids loading of 0.070 lbs VSS/ft^3-day and an air loading rate of 0.060 cfm/ft^3. Cationic polymer was used for conditioning. The data collected is included in Figures 44 and 45. Additionally, various mixtures of aerobically digested waste activated sludge and anaerobically digested primary sludge were combined in a 2000 gallon holding tank and then centrifuged. The waste activated to primary sludge ratios investigated were 70% WAS - 30% primary and 50% WAS - 50% primary. The data from these evaluations are also included in Figures 44 and 45.

For all test runs, the centrifuge was operated at a bowl speed of 2070 rpm (1200 g's) and a relative scroll speed of 19 rpm.

A maximum pool depth was maintained for the straight waste activated sludge and the 70% WAS - 30% primary mixture while three pool depths were evaluated on the 50% WAS - 50% primary sludge mixture. As seen in Figure 44, cakes ranging from 8% to 10.5% TS were obtained on aerobically digested waste activated sludge with the addition of 9 to 31 lb/ton of polymer. Centrate quality deteriorated rapidly at polymer dosages less than 20 lb/ton and a minimum dosage of 17 lb/ton was required for effluent suspended solids concentrations to be 1500 mg/l or less (Figure 45). When comparing these two curves (Figures 44 and 45) with the two developed when the digester was operated at an eight-day detention time (Figures 42 and 43), it is readily seen that the sludge aerobically digested for eight days exhibited better dewatering characteristics than the sludge aerobically digested for 12.7 days. This same relationship existed in the work done by Parker et al[8] whereby the effects of aerobic digestion detention time on waste activated sludge filterability were investigated. They showed that "aeration produced an initial improvement in sludge filterability, with a maximum improvement in 4 to 6 days. However, greater periods of aeration caused an increase in filtration time with the result that after two weeks of digestion the sludge was almost as difficult to dewater as it was before aerobic digestion."

The effects of combining anaerobically digested primary sludge with the aerobically digested waste activated sludge are also presented in Figures 44 and

45. With the centrifuge operating at a feedrate of 30 gpm and the maximum pool depth setting, the resultant cake solids increased to 11% - 12% TS on 70:30 combined digested sludges, and with the 50:50 combined digested sludges, the cake solids increased to 13% - 15% TS. Suspended solids recoveries also increased with increases in the amount of digested primary sludge. The effects of pool depth on the 50:50 combined digested sludges are also shown in these two figures. Cake solids increased slightly and suspended solids recovery decreased slightly as the liquid pool depth was varied from maximum to minimum. A wider range of pool depths and its subsequent effects on cake solids and recovery could not be investigated because this particular centrifuge does not have the capability of varying pool depths beyond those investigated.

As the anaerobic digester was gradually converted to WAS feed, various ratios of waste activated and primary sludge have been digested. These blend ratios ranged from 23% WAS - 77% primary to 100% WAS - 0% primary. Centrifuge data were collected on the 23:77, 31:69, 43:57, and 70:30 digested blends. Centrifugation of anaerobically digested waste activated sludge (100% WAS - 0% primary) was also evaluated. In addition, centrifuge data were collected on various combinations of digested primary and digested waste activated sludge and various combinations of digested primary plus 70:30 digested blend.

As a matter of interest, the average performance and operating parameters maintained in the primary sludge digester during periods when digested primary sludge was combined with either a digested blend or digested waste activated sludge are presented in Table XXVI.

Data collected on scroll centrifuge dewatering of a 23:77% digested blend is presented in Figures 46 and 47. A 32" x 100" tapered bowl scroll centrifuge was used to dewater this particular sludge mixture from the Saugus-Newhall WRP and the Valencia WRP. The Valencia WRP utilizes secondary digesters for separation; hence, secondary digester subnatant (SDS) and primary digester supernatant (PDS) were subjected to dewatering through the 32" x 100" scroll centrifuge.

The maximum cake solids obtained for the Saugus-Newhall WRP sludge approximated 11% TS while cake solids of 15% TS were obtained on both sludge sources from the Valencia WRP. Centrate suspended solids of 1500 mg/l or less were consistently obtained at polymer dosages in excess of 13 lb/ton as shown in Figure 47. It should be noted that the 32" x 100" scroll centrifuge had been optimized with regard to pool depth prior to the collection of this data. Resultant cake solids did not vary significantly with variations in the liquid pool depth but the desire to obtain a relatively clear centrate (1500 mg/l SS) required that the unit be operated under maximum pool depth conditions. Upon the recommendation of the manufacturer, the unit was operated at a bowl speed of 1280 rpm (750 g's) and a relative scroll speed of 16.5 rpm.

A digester at the Valencia WRP was fed a blend of 31 percent WAS - 69 percent primary sludge. The data from dewatering the 31:69 digested blend on a 14" x 48" scroll centrifuge are presented in Figures 48 and 49. The cake solids ranged from 12 percent to 14 percent TS at polymer dosages of 6 to 13 lb/ton, but a minimum dosage of 8 lb/ton of polymer was required to produce a centrate with less than 1500 mg/l SS.

Dewatering experiments on the 43:57 digested blend from the Saugus-Newhall WRP were conducted utilizing the 20" x 62" tapered bowl scroll centrifuge. The centrifuge was operated at constant bowl and relative scroll speeds of 2070 rpm (1200 g's) and 19 rpm, respectively, while the pool depth was varied from its maximum depth to its mid-depth setting.

As seen in Figure 50, the effect(s) of pool depth on cake solids was very slight over the range of polymer dosages investigated (15 - 33 lb/ton). At the maximum pool depth setting, cake solids of approximately 14% TS were obtained whereas cake solids of approximately 15% TS were obtained at the midpool depth setting. The variations in pool depth had a more pronounced effect on centrate solids and solids recovery as shown in Figure 51. Under maximum pool depth conditions, centrate solids of less than 500 mg/l were obtained over the polymer range investigated whereas at a polymer dosage of 22 lb/ton and midpool depth conditions the centrate stream contained 10,000 mg/l of suspended solids. Polymer dosages less than 15 lb/ton were not investigated because problems developed within the centrifuge and the investigation had to be terminated.

Dewatering data collected on the 70:30 digested blend utilizing the 20" x 62" scroll centrifuge are included in Figures 52 through 55. In addition to dewatering this digested blend, combining with various amounts of digested primary sludge prior to centrifugation were also investigated. This data is also included in Figures 52 through 55.

Figures 52 and 53 represent the data collected while the centrifuge was operated at a bowl speed of 2070 rpm (1200 g's), a relative scroll speed of 19 rpm and a maximum pool depth. With the addition of 8 to 17 lb/ton of cationic polymer, the combined digested sludge dewatered to 11 - 13% TS and a minimum of 17 lb/ton of polymer was required for the centrate to contain less than 1500 mg/l of SS. As the amount of digested primary sludge was increased in the blended mixtures, the dryness of the resultant cakes was observed to also increase while the polymer requirements to achieve 95% solids recovery decreased. It should be noted that as the percentage of digested primary sludge increases and an increase in handleability is realized, the final dryness of the resultant sludge mixtures are equivalent to that which would be achieved if the sludges are dewatered separately and then combined.

Figures 54 and 55 summarize the data collected while the centrifuge was operated at its midpool depth. The bowl and relative scroll speeds were respectively 2070 rpm and 19 rpm. Resultant cake solids for the combined digested sludge varied from 12 - 15% TS with the addition of 11 to 23 lb/ton of cationic polymer.

A minimum polymer requirement of 17 lb/ton was again required to produce a centrate containing less than 1500 mg/l of SS but at the midpool depth setting, the deterioration in centrate quality with decreasing polymer dosages was more rapid than under maximum pool depth conditions. Again, as the amount of digested primary sludge increased, the handleability of the resultant sludge mixtures also increased, while the degree of dewatering was the same as had the sludges been blended after dewatering.

The successful digestion of 100% waste activated sludge enabled dewatering data to be collected, utilizing the 20" x 62" scroll centrifuge. The anaerobically digested waste activated sludge exhibited poor dewatering characteristics and required a minimum of 35 lb/ton of cationic polymer in order for the centrate flow to contain 1500 mg/l or less SS. Three different cationic polymers were used for pre-conditioning, but in each case a minimum dosage of 35 lb/ton was required and, as seen in Figure 56, the maximum cake dryness obtained approximated 12% TS. The physical appearance or characteristics of the cakes obtained on the digested waste activated sludge suggests that little or no free moisture exists.

In fact, the discharge cakes for this particular digested sludge are plastic in nature and appear as dry as 20 - 25% TS cakes obtainable upon centrifugation of digested primary sludge.

Various mixtures of anaerobically digested primary and anaerobically digested waste activated sludge were combined (on a solids basis) in a 2,000 gallon holding tank prior to dewatering through the 20" x 62" scroll centrifuge. During these dewatering experiments, the centrifuge operating parameters were held constant with a bowl speed of 2,070 rpm (1200 g's), a relative scroll speed of 19 rpm, and a maximum pool depth setting, while the feed rate was maintained at 30 gpm. The data collected is included in Figures 56 and 57. As presented in the plot of cake solids versus polymer dosage, cake dryness increased with increasing amounts of digested primary sludge and, as seen in Figure 57, the polymer requirements necessary for the centrate to contain less than 1,500 mg/l SS decreased significantly with increasing quantities of digested primary sludge.

The effects of blending various ratios of digested primary sludge and digested waste activated sludge on resultant centrifuge cake solids, and polymer requirements to achieve 1,500 mg/l SS or less in the centrate, are more clearly defined in Figure 58. A linear relationship exists between maximum obtainable cake solids and the ratio of anaerobically digested waste activated to digested primary sludge. The 100% digested primary sludge dewatered to 22% TS while cake solids decreased linearly to 12% TS upon centrifugation of 100% digested waste activated sludge. As the quantity of digested primary sludge increased, the dryness of the resultant centrifuge discharge solids also increased; but not beyond that which would be achieved if the dewatered sludges are blended following separate centrifugation. Polymer requirements increased linearly with increasing amounts of waste activated sludge up to a waste activated to primary sludge ratio of 40:60%. From 40% WAS - 60% primary to a 70% WAS - 30% primary sludge mixture, the polymer requirements for the centrate to contain 1,500 mg/l or less of suspended solids remained at 18 lb/ton, then increased linearly to 35 lb/ton for 100% digested waste activated sludge.

Combined Digestion and Dewatering vs. Separate Digestion and Dewatering--

Consideration should be given to the dewaterability of digested blends as compared to separate digestion followed by combining and dewatering. It might be assumed that there is no difference in dewaterability if the sludges are blended prior to digestion and dewatering, or combined and dewatered subsequent to separate digestion. Data contrary to this assumption are presented in Figures 59 and 60. The 70:30 and 43:57 digested blends were dewatered in the 20" x 62" tapered bowl, scroll centrifuge. For comparison purposes, the same waste activated to primary sludge ratios were combined following separate anaerobic digestion and dewatered through the same unit. Although dewatering data were collected on various sludge blends, the two ratios listed above are the only ones that can be isolated for direct comparison, because the 20" x 62" scroll centrifuge was the only unit on site when anaerobically digested waste activated sludge was available. In both cases, separate digestion followed by blending and centrifugation produced dryer discharge cakes than the flow scheme incorporating combined digestion followed by centrifugation. Polymer requirements to produce centrate SS of 1,500 mg/l or less approximated each other for the two sludges and were in the range of 15-18 lb/ton. At these polymer dosages, cake solids for the 70% WAS - 30% primary mixtures were 12 - 13% TS and 14 - 15% TS, respectively, for digested blends and combined digested sludges. The dewatering benefits realized by separate digestion followed by combining and centrifugation were more apparent on the 40% WAS - 60% primary sludge blends. As shown in Figure 59, centrifuge discharge solids approximated 18 - 20% TS for separately digested combined sludge while cake solids of 14 - 15% TS were obtained on the digested blend.

When comparing the centrifuge data collected on aerobically digested waste activated sludge (Figures 44 and 45) with that collected on anaerobically digested waste activated sludge at the Saugus-Newhall WRP (Figures 56 and 57), it can be seen that the dewaterability of the aerobically and anaerobically digested sludge approximate each other. To further substantiate that aerobic digestion of sludges does not enhance dewaterability, aerobically digested sludges from two other wastewater treatment plants were hauled to the Saugus-Newhall WRP and dewatered via centrifugation. Each of the sludges consisted of approximately 70% primary - 30% WAS that has been aerobically digested for approximately fifteen (15) days. In each case, the dewaterability was equivalent to that of combined anaerobically digested sludge. For centrifugation cake solids were in the range of 10% - 13% TS and approximately 15 lb/ton of polymer were required for a centrate suspended solids concentration of less than 1500 mg/l.

The scroll centrifuge evaluations at the JWPCP were conducted on an 18" x 54" machine. This unit's dewatering data on mesophilically digested oxygen WAS are presented in Figures 61 and 62. Because of the small daily volume of WAS produced, relatively few centrifuge runs were conducted on digested oxygen sludge alone. Flow rates of 10 and 15 gpm were evaluated while the unit was operated at its maximum pool depth (3.94"). A constant bowl speed of 1550 rpm, corresponding to an acceleration force of 650 g's, was maintained while the differential scroll speed was varied from 1.0 to 3.3 rpm. Cationic polymer addition ranged from 8.5 to 15 lb/ton and, as seen in Figure 61, within this range of dosages the discharge solids were consistently in the range of 15 to 17 percent TS. Cake solids increased with increasing chemical dosage and decreasing differential scroll speeds while suspended solids removal decreased at the lower relative scroll speeds. As seen in Figure 62, SS recovery varied from 60 to 95 percent over the range of parameters investigated. With the exception of the run made at a differential scroll speed of 1.0 rpm, polymer requirements varied from 10 to 13 lb/ton for the centrate stream to contain 1500 to 2000 mg/l SS.

A review of centrifuge data collected at the Saugus-Newhall WRP on digested WAS indicated that either the JWPCP digested oxygen WAS has better dewatering characteristics, or the 18" x 54" centrifuge was more suited for dewatering these sludges than the 20" x 62" scroll centrifuge. To resolve this question, 5000 gallons of digested WAS from the Saugus-Newhall WRP were hauled to the JWPCP research site and dewatered via the 18" x 54" scroll centrifuge. The results from this evaluation are presented in Figures 63 and 64. With the addition of polymer only, cake solids approximated 10 percent TS while previous testing with the 20" x 62" scroll centrifuge at the Saugus-Newhall WRP yield cake solids from 10 to 11 percent TS. The fact that both centrifuges dewatered the digested WAS from the Saugus-Newhall WRP to approximately 10 percent TS and the digested oxygen waste activated sludge from the JWPCP was dewatered to 15 to 17 percent TS (Figure 61) indicates that the oxygen WAS exhibits better dewatering characteristics.

A series of tests were also conducted on the digested WAS from the Saugus-Newhall WRP with the addition of 100 lb/ton of alum and 9-to-15 lb/ton of polymer. These results are included in Figures 63 and 64 and indicate that alum addition substantially enhanced the dewaterability of the sludge. Based on these few runs, it was decided to evaluate alum and polymer addition on the JWPCP digested oxygen waste sludge. The results from this evaluation are presented in Figures 65 and 66 and indicate that alum addition decreased the dewatering characteristics of this particular sludge. The results obtained on the 18" x 54" scroll centrifuge for the sludges generated at the JWPCP and the Saugus-Newhall WRP (Figures 14 through 17) are indicative of the unpredictable responses encountered when dewatering sludges generated at different treatment plants. Prior to terminating the mesophilic digestion studies, various combinations of the JWPCP digested primary sludge and digested oxygen WAS were dewatered through the 18" x 54" scroll centrifuge. The data presented in Figures 67 and 68 show the effects of increasing the ratio of digested WAS on cake solids, centrate quality, and chemical requirements.

To more clearly show the above effects, Figure 69 is a plot of the required chemical dosage necessary to achieve 95 percent SS recovery for the various WAS to primary sludge ratios. Also included in Figure 69 is the resultant cake solids obtained at these polymer dosages. When comparing this data with that collected at the Saugus-Newhall WRP (Figure 58), it becomes evident the digested WAS generated at the JWPCP dewatered more readily than the sludge encountered at the Saugus-Newhall WRP.

Basket Centrifuge--

A number of basket centrifuges were evaluated during this study. All were similar in design, and all had 48" diameter bowls, so no distinction will be made in this report between the machines.

Only a very limited basket centrifuge study was conducted on aerobically-digested sludge because of equipment scheduling conflicts. During November 1974, when the aerobic digester was operated at a hydraulic detention time of eight days, a volatile solids loading of 0.081 lb VSS/ft^3 day, and an air loading of 0.043 cfm/ft^3, one set of centrifuge operating conditions was evaluated for dewatering the aerobic digester mixed liquor. Rotating at 1,380 rpm (1300 g's), the centrifuge was fed aerobically digested waste activated sludge with a suspended solids concentration of 0.90 percent at a feedrate of 25 gpm. Cationic polymer was added in the range of 5 to 18 lb/ton. The results from this brief evaluation are shown in Figure 70. The maximum cake solids obtained was approximately 10% with the addition of 18 lb/ton of polymer. Suspended solids recovery was 99% or better for each of the chemical dosages evaluated.

No other dewatering equipment was available for evaluation when the aerobic digester was operated under the same parameters. Some comparison may be made between the basket centrifuge data and the scroll centrifuge data from December 1974 through March 1975 (Figures 42 and 43). The digestion detention time and solids loadings were comparable during the two periods, but the air rate was almost 50 percent higher during the period when the scroll centrifuge data were obtained. The two types of centrifuges produced comparable cakes at 8 to 10 percent TS, but the basket centrifuge obtained much better SS recoveries. It is not possible to estimate, however, what effects the change in air rate had in the dewaterability of the digested WAS.

The data collected at the Saugus-Newhall WRP utilizing a 48" basket centrifuge to dewater the 23:77 digested blend are presented in Figures 71 and 72. The maximum cake solids obtained approximated 11% TS while the centrate suspended solids were less than 1500 mg/l at polymer dosages in excess of 10 lb/ton. Visual observations while loading the unit in excess of 400 lbs/hr indicated the centrate was of extremely poor quality and for this reason the maximum applied solids loading was 400 lbs/hr. Lower solids loadings were not investigated because the feed pump to the centrifuge had a minimum capacity of 35 gpm. The cake solids and SS recoveries of the basket centrifuge and the scroll centrifuge (Figures 46 and 47) were equivalent on this digested blend.

The data collected on dewatering the 31:69 digested blend from the Valencia WRP are presented in Figures 73 and 74. The Valencia WRP employs two-stage digestion and both primary digester supernatant and secondary digester subnatant were dewatered on the basket centrifuge. With cationic polymer added in the range of 5 to 30 lb/ton, the maximum cake solids obtained approximated 13-15% TS for both sludge sources. Centrate suspended solids were consistently less than 1500 mg/l over the polymer range investigated but higher solids recoveries were recorded while feeding secondary digester subnatant because of a higher solids concentration in this feed source.

These data are comparable to the scroll centrifuge except that the 1500 mg/l centrate SS criterion was met at a lower polymer dosage with the basket centrifuge.

At the JWPCP, basket centrifugation of the digested oxygen WAS with the addition of 5 to 15 lb/ton of cationic polymer yielded total cake solids from 5 to 9 percent TS with 90 to 95 percent recovery of the applied SS. The 48" basket was operated at a bowl speed of 1380 rpm and hydraulically loaded at 50 to 35 gpm. The effect of polymer dosage and hydraulic loading rate on discharge solids and suspended solids recovery are shown in Figures 75 and 76, respectively. The cake solids data shown represents a composite of the basket solids and includes approximately 2.5 cubic feet of skimmings and 13.5 cubic feet of plowed solids. It should be noted that the basket run times approximated 8 minutes at the 35 gpm feed rate and 5 minutes for feed rates of 50 gpm. These short run times in conjunction with the time required to decelerate and knife out the basket contents yield an effective flow rate of approximately 25 gpm.

Further indication that the digested oxygen WAS from the JWPCP was less difficult to dewater than the WAS from the Saugus-Newhall WRP can be found in the basket centrifuge data. The same 48" basket centrifuge was used at the Saugus-Newhall WRP to dewater the 23:77 digested blends. With the addition of 5 to 15 lb/ton of cationic polymer, total cake discharge solids varied from 8 to 11 percent TS while suspended solids recoveries varied from 90 to 97 percent. Realizing that this sludge consisted of only 23 percent waste activated sludge and responded almost identically to the 100 percent oxygen waste activated sludge from the JWPCP, it becomes more evident that the biological sludge generated at the JWPCP is less difficult to dewater.

Process Selection

Centrifugation seems to be the best process for dewatering digested WAS. The filtration options required extremely high chemical dosages. The chemicals involved are comparable in expense, but considerable materials handling problems will be encountered. Basket centrifuges require less chemical than the scroll centrifuges, but, since they operate in a batch mode, basket centrifuges require more operator attention. These factors will be discussed in more detail in the cost analysis section of this report.

THERMAL CONDITIONING AND DEWATERING

Thermal conditioning of WAS was evaluated as an alternative to digestion and chemical conditioning. Thermal treatment is basically a continuous pressure cooking process. The excess sludge is heated under pressure so that the proteinaceous material composing the cell walls is hydrolyzed and the bound water is released, thereby permitting the sludge to dewater more readily. Additionally, under normal conditions of thermal conditioning, all pathogenic organisms should be destroyed due to the high temperatures and detention times maintained. Typically, the sludge temperature is elevated to 350°F - 400°F, the pressure is raised to 300-400 psi and the retention time is between 20 and 40 minutes.

A 40' mobile trailer-mounted continuous flow thermal sludge conditioning pilot plant was operated at the Saugus-Newhall WRP and at the JWPCP. A schematic of the process is shown in Figure 77. The feed sludge is passed through a grinder to reduce all particles to 1/4" dimensions before pumping. Sludge (with air when operated under the wet oxidation mode of operation) is then pumped to the system where it is passed through heat exchangers and brought to the initiating reaction temperature as it enters the reactor. Oxidation takes place in the reactor and the oxidized products leaving the reactor are cooled in the heat exchangers against the entering cold sludge. The oxidized liquid and remaining suspended solids are released to a decant tank for separation and compaction while the gases are released through a pressure control valve to an odor control system. The overflow from the thickener or decant tank may be returned to the head end of the treatment plant while the thickened subnatant is pumped to subsequent dewatering units. For start-up and whenever the process is not thermally self-sustaining, heat is added from an outside source such as the steam generator incorporated in the pilot trailer.

Prior to dewatering, the thermally conditioned sludge flowed to a decant tank. The decant tank installed on the mobile trailer was oversized and would have yielded detention times of 4.4 to 6.5 hours had the unit been operated continuously. In actuality, the thermal conditioning unit was operated in a batch manner and after a sufficient amount of sludge had accumulated in the decant tank the thermal conditioning process was aborted and the decant tank underflow dewatered. Because of the intermittent mode of operation, the actual solids retention time in the decant tank approximated 24 to 48 hours prior to dewatering. The manufacturer of the thermal unit agreed that the decant tank detention time was excessive and indicated that under normal practice detention times of 11 to 24 hours are common.

Three modes of thermal conditioning were investigated over the course of this evaluation. These were, low pressure wet oxidation (LPO), intermediate pressure wet oxidation (IPO), and heat treatment (HT). Under the low pressure wet oxidation mode of operation, the thickened waste activated sludge was reacted with air while temperatures of 380°F to 400°F and a pressure of 400 psig were maintained in the reactor. Under the intermediate pressure wet oxidation mode of operation, the sludge was again reacted with air while a temperature of 450°F and a reactor pressure of 500 psig was maintained. Heat treatment proceeded with the same operating conditions as LPO with the exception that no air was introduced into the reactor. Various sludge feedrates and, consequently, different reactor detention periods or "cooking times" were also investigated. The results obtained in the operation of the unit and the subsequent dewatering of the sludge are discussed below.

Low Pressure Wet Oxidation

Under the LPO mode of operation the sludge feedrate to the unit was varied from 3 to 6 gpm corresponding to reactor detention periods between 24 and 48 minutes. The reactor pressure was held constant at 400 psig while reactor temperatures of 380°F and 400°F were investigated. Composite samples of the feed sludge, oxidized sludge, herein referred to as "heatrate" and thickener overflow were collected and analyzed at the treatment plant laboratories.

Summaries of the operating conditions, solids data, and COD and BOD data from the Saugus-Newhall Studies are presented in Tables XXVII, XXVIII, and XXIX, respectively. Eight separate runs were made while incorporating the LPO mode of thermal conditioning and these are designated SNLPO 1 through SNLPO 8. The first seven were made on a 100% waste activated sludge feed source while a combination of 77% waste activated and 23% primary sludge was fed to the reactors for run number 8.

Thickened waste activated sludge feed solids varied in concentration from 2.16% TS to 3.28% TS for runs 1 through 7, and volatile solids destruction ranged from 26 to 49 percent with no apparent correlation to reactor temperature or cooking time. The greatest reduction of solids was achieved at a cooking time of 29 minutes with reactor temperatures of 380°F - 400°F. A volatile solids destruction of 55% was achieved on the combined feed source (77% WAS - 23% primary) at a cooking time of 29 minutes and a reactor temperature of 390°F.

Regardless of the cooking time or pressure investigated, the low pressure wet oxidation process effected an increase in dissolved solids, a decrease in total and suspended solids, and an increase in soluble organics. In fact, as seen in Table XXIX, soluble COD increased from 38 to 63 times and soluble BOD increased from 11 to 32 times as the waste activated sludge was thermally conditioned. The thermal solubilization of organics is one of the major disadvantages to this type of stabilization because of the increased organic load on any treatment works when supernatant liquor is returned to the inlet works of the treatment system. Another problem encountered with thermal conditioning is odor production. For lack of a more descriptive name, the odors generated from the thermal unit were classified as having the characteristics of "burnt coffee" or "burnt plastic", and even with the catalytic combustion unit for odor control the odors were distinctive and highly offensive.

Summaries of the operating conditions, solids data, COD data, and BOD data collected at the JWPCP are presented in Tables XXX, XXXI, and XXXII. Six separate runs were made under LPO conditions and are designated JLPO 1 through JLPO 6. The third low pressure oxidation run (JLPO 3) was made on thermophilically digested oxygen waste activated sludge while the remaining five runs utilized thickened oxygen WAS or a combination of thickened oxygen plus air WAS.

Solubilization of solids and organics are evidenced by an increase from 2000 mg/l to 12,000 mg/l in the dissolved solids concentration and an average increase from 1400 mg/l to 14,000 mg/l in coluble COD concentrations as the oxygen waste activated sludge was subjected to low pressure wet oxidation. The degree of solubilization could not be predicted by the process operating parameters and no correlations exist between solubilization obtained and operating temperatures or reactor detention times maintained. Thermal runs JLPO 1 and JLPO 2 were made at an operating temperature of 380°F and reactor detention times of 29 and 41 minutes, respectively. The longer retention time should yield a higher degree of solubilization but the data shows the reverse. Conversely, runs JLPO 4, 5, and 6 were conducted at 400°F and detention times of 41, 29, and 48 minutes, respectively, and the most solubilization occurred at 41 minutes of reactor detention time (JLPO 4). These same inconsistencies were observed at the Saugus-Newhall WRP and indicated that problems will exist in predicting process results for a full scale installation. A more substantial indication of the unpredictable performance of the thermal conditioning unit is given by the COD oxidation values. For each of the LPO conditions the theoretical percent of COD oxidation can be predicted based on the amount of air introduced into the reactor. Table XXXIII summarizes the theoretical and actual oxidation values obtained for each of the low pressure oxidation runs. As presented, the actual COD oxidation varied from 65 percent below to 93 percent above the theoretical value with no apparent correlation to system operating parameters. For those runs in which the measured reductions were greater than the theoretical values, it is conjectured that dilution of the processed sludge with steam occurred within the system reactor. Steam is injected into the reactor whenever the system is not thermally self-sustained and depending on the amount added the degree of dilution will vary. For those runs with measured reductions, lower than the theoretical estimates, it can be concluded that the oxygen transfer efficiency was low and indicative of plugging or scaling conditions within the reactor. In view of the above concerns; namely, steam dilution; the remaining data presented in Tables XXXI and XXXII must be reviewed with caution because the exact dilution factors are not known and the data does not reflect corrections for steam dilution.

Under normal thermal conditioning operation, complete pathogenic organism destruction should be accomplished due to the high temperatures and detention times obtained. The measured reductions in total and fecal coliforms, as presented in Table XXXIV, were erratic during these low pressure wet oxidation studies. The reason(s) for the high coliform counts in the thermally treated sludge are not thought to be related to either sampling or laboratory technique. As suggested by the manufacturer of the thermal conditioning unit, the high coliform counts are thought to be the result of regrowth in the pipes connecting the heat exchanger outlet and the decant tank. The sample tap for the heatrate stream is located downstream of the reactor and heat exchangers, but upstream of the decant tank, and the manufacturer contends that regrowth is occurring within this line. The manufacturer suggested that samples of the heatrate be taken on the reactor discharge line prior to flowing through the heat exchangers and subsequently to the decant tank. Due to the high temperature and pressure of the reactor effluent stream, a special sampling container had to be manufactured and at the time of this writing had yet to be supplied. The problems surrounding the coliform kill data are indeed perplexing but not uncommon to the host of other unexplained phenomena surrounding the thermal treatment pilot studies.

Dewatering of LPO conditioned WAS was accomplished on an 8.4 ft^2 filter press and a 3' x 1' rotary drum vacuum filter. Data collected on dewatering via the vacuum filter are presented in Figures 78 through 82. Selection of the cloth media was made by the manufacturer of the thermal conditioning unit and three cloths accompanied the vacuum filter. Initially, data were collected on all three cloths but based on process performance data collected at the Saugus-Newhall WRP, it was decided to use one cloth for the filtration studies. A tightly woven nylon fabric served as the filtration media and the data presented herein were all collected on the same cloth.

Vacuum filtration data on thickened waste activated sludge that had been subjected to LPO conditioning at 380°F and three different detention periods are presented in Figure 78. The sludge conditioned for 36.4 minutes yielded higher filter yields than that conditioned for 26.5 minutes. These two sludge sources were thickened to 8.3% SS and 8.5% SS, respectively, for the 36.4 and 26.5 detention periods prior to dewatering. Filter yields increased with lower cycle periods and the cake solids data exhibited a random effect with various drum cycle times. Cake solids were consistently greater than 31% TS and it should be noted that no chemicals were added for preconditioning prior to dewatering and the captured solids exhibited good discharge characteristics. The sludge conditioned for 29.2 minutes at 380°F yielded the highest filter yields at all of the cycle times investigated but the sludge was thickened to 12.61% SS prior to dewatering and it is believed that the yields increased because of the high solids loading and not the result of a 29.2 minute detention period. Suspended solids recovery through the vacuum filter for each of these runs exceeded 97% while the overall suspended solids removal through the thermal unit and vacuum filter consistently exceeded 93%.

The effects of varying reactor temperature from 380°F to 400°F and maintaining a constant reactor detention period on cake solids and filter yields is presented in Figures 80 and 81. With a reactor detention time of 36.4 minutes, no significant changes in dewaterability were noted by increasing the reactor temperature from 380°F to 400°F. At a reactor detention time of 29.2 minutes, the sludge exhibited somewhat better dewaterability when the reactor was operated at 380°F, but the difference is probably attributed to the higher feed solids concentration to the filter and not the variation in temperature.

The effects of blending primary sludge wizh thickened waste activated sludge (23% RAW - 77% WAS) prior to low pressure wet oxidation on dewaterability is presented in Figure 82. Only one run was made on thermal conditioning of a combined sludge. The reactor was operated at a temperature of 380°F, a pressure of 400 psig and the sludge was retained in the reactor for 29.2 minutes. Following LPO conditioning and decant thickening, the sludge was fed to the vacuum filter. As seen in Figure 82, cake solids increased slightly and filter yields decreased for the combined sludge as compared to thermally conditioned waste activated sludge.

At the JWPCP, vacuum filtration of LPO conditioned oxygen WAS yielded cake solids varying from 31 to 34 percent TS and filter yields of 3.0 to 6.6 lb/hr-ft^2. Drum cycle times were varied from 2 to 8-1/2 minutes while applied vacuums varied from 15 to 20 inches Hg. The resultant suspended solids removal efficiencies were erratic and varied from 61 to 95 percent and, as seen in Tables XXIII A and B, the discharge characteristics were consistently good, indicative of self-discharging and minimal media cleaning required. The vacuum filter was not available for runs JLPO 1, JLPO 2, and JLPO 3.

Data collected on filter pressing of LPO conditioned Saugus-Newhall WAS was disappointing with regard to filter yields. Two nylon cloths (recommended and supplied by the manufacturer) were employed as filtration media and pressures varying from 102 to 120 psi were applied to the press. Filter yields averaged 0.50 lb/hr-ft^2 while cake solids varied from 36 to 56% TS. It should be noted that the particular filter press used during these studies could sustain a maximum pressure of only 120 psi while other manufacturers can supply presses constructed to withstand pressures in excess of 220 psi. This latter type of press has been used in previous Districts' programs but was not available during the time the thermal conditioning unit was on site. A summary of the data collected on pressure filtration of LPO conditioned Saugus-Newhall WRP is presented in Table XXXVI.

Data collected on the pressure filter for each of the low pressure oxidation runs at the JWPCP are summarized in Tables XXXVII A and B. For each of these runs the filter was operated at a pressure of 100 psig and utilized a cloth previously used in conjunction with the Saugus-Newhall WRP thermal studies.

JLPO 1 and JLPO 2 sludges were conditioned at 380°F for respective periods of 29 and 41 minutes prior to dewatering. The cake solids and filter yields for the JLPO 2 were higher than those obtained for a 29-minute thermal conditioning period, indicating a slight advantage to increased conditioning time. The consistency of the filter cakes for these two sets of runs were listed as good, indicating the cakes were consistently firm throughout. The discharge characteristics were poor, indicating that the cakes did not fall by gravity when the press was opened, and considerable cleaning of the media was required.

Pressure filtration of LPO conditioned digested oxygen WAS (JLPO 3) produced cake solids in the range of 29 to 32 percent TS that were, for the most part, unconveyable. The consistency of the discharge solids were poor, indicating that an "egg shell" effect was observed. The portion of solids nearest the filter plates were firm but when these plates were disengaged the cake would split along the centerline and discharge a liquid center. The thermal conditioning operating parameters were identical for runs LPO 1 and LPO 3 and the resultant filter press data collected on these two sets of thermal runs is indicative of the adverse effects that digestion has on dewaterability of thermally conditioned WAS.

The dewatering advantages afforded by LPO conditioning at 400°F as opposed to a reactor temperature of 380°F are evident by the pressure filtration data shown in Table XXXVII B. Thermal runs LPO 4 and 5 were conducted at a temperature of 400°F and the resultant filter press cakes were consistently in excess of 47 percent TS while the filter yields (lb/hr-ft^2) were higher than those conducted on JLPO 1 and 2 sludges. For each of these experiments the cake consistency was good while the discharge characteristics were poor. The exception was for the run which employed an application of approximately 10 lb/ton of diatomaceous precoat prior to the filtration sequence. For this rin, the cake discharge was fair, indicating it did not discharge by gravity; but it left the filter media clean, requiring minimal cleaning.

Thermal run JLPO 6 employed a temperature of 400°F and a reactor detention time of 48 minutes. This was the longest thermal conditioning time employed and, as seen in Table XXXVII B, a slight increase in dewaterability was observed. Cake solids approximated 50 percent TS with good consistency and even without precoating had fair discharge characteristics.

Heat Treatment

Under the heat treatment mode of operation, reactor temperatures of 380°F and 400°F were maintained while the reactor operating pressure was held constant at 400 psig. The sludge feed rate was varied from 3.5 to 6.2 gpm, corresponding to detention times of 41 to 24 minutes. Tables XXXVIII and XXXIX summarize the operating conditions maintained and the sludge source conditioned for each of the eight heat treatment runs.

The solids, BOD, and COD data from the Saugus-Newhall studies are presented in Tables XL and XLI. For each of the three runs, thickened waste activated sludge served as the feed source with feed solids concentrations ranging from 2.36 to 2.91% TS. Volatile solids reduction varied from 24 to 41%, again with no apparent correlation to reactor temperature or detention time. As was the case with LPO conditioning, heat treatment affected an increase in dissolved solids and soluble organics and a decrease in total and suspended solids. As seen in Table XLI, thermal solubilization of organics is evident by an increase in soluble COD of 17 to 65 times and an increase in soluble BOD of 8 times. During these heat treatment studies, offensive "burnt coffee" odors were again present and easily detectable.

Thermal run JHT 1 was made on digested oxygen WAS while the remaining heat treatment runs at the JWPCP utilized undigested oxygen or oxygen plus air WAS. Summaries of the solids data and COD and BOD data are presented in Tables XLII and XLIII, respectively. Solubilization of solids and organics are evidenced by an increase from 200 mg/l to 12,500 mg/l in dissolved solids and from 1600 mg/l to 15,300 mg/l in the soluble COD concentrations as the oxygen waste activated sludge was heat treated.

It is interesting to note that under heat treatment conditions, little, if any, reduction in total COD should occur because of the omission of air addition into the thermal reactor. As seen in Table XLIII, measured reductions in total COD varied from 9 to 35 percent, indicating that steam dilution is occurring but the exact amount is not known. The problems of coliform reduction encountered under LPO conditions were again manifested during the heat treatment studies, and as shown in Table XLIV, the measured reductions were extremely erratic.

The underflow sludge from the thickener was dewatered via the vacuum filter and filter press. The data collected on vacuum filtration of the Saugus-Newhall sludge are presented in Table XLV. A comparison of the data collected on filtration of heat treated sludge at 380°F and a detention time of 29.2 minutes (SNHT1) with that collected under LPO conditions at the same temperature and detention time (Figure 78) indicates that LPO conditioning slightly enhances sludge dewaterability. Under LPO conditions cake solids approximated 32% TS and a filter yield of 3.5 $lb/hr\text{-}ft^2$ was obtained. For heat treatment conditioning, the cake solids approximated 30% TS while the filter yield decreased to 0.72 $lb/hr\text{-}ft^2$. Suspended solids removal through the filter approximated 97% while the net removal through the thermal conditioning unit and filter approximated 93%.

Vacuum filtration of the heat treated oxygen WAS, however, was very comparable to the results obtained under LPO conditioning. As seen in Table XLVI, cake solids varied from 33 to 37 percent TS while filter yields were measured at 2.3 to 5.9 $lb/hr\text{-}ft^2$. Suspended solids recoveries were consistently in excess of 91 percent. In this regard it should be noted that prior to the heat treatment studies the vacuum filter manufacturer supplied a new filter cloth. This new cloth was identical to the one previously used in conjunction with the LPO studies but employed a rubber gasket to ensure a good seal between the filter drum and filtration media. During the LPO experiments at the JWPCP an adequate seal may not have developed; hence, solids may have passed through the outer edges of the drum instead of through the cloth. The suspended solids data presented on vacuum filtration of LPO conditioned sludge (Table XXXV) may therefore be erroneous, but the cake solids and yield data are thought to be reliable.

Pressure filtration was conducted on heat treated Saugus-Newhall WAS conditioned at 400°F, 400 psig for 29.2 minutes. These results are presented in Table XLVII. Cake solids approximated 45% TS but, as experienced previously with LPO conditioned Saugus-Newhall sludge, the filter yield was extremely low at a value of 0.27 $lb/hr\text{-}ft^2$. When comparing this data with that collected under the same operating conditions and the LPO mode of treatment (Table XXXVI), it is again evident that LPO conditioning yields a sludge slightly easier to dewater for the Saugus-Newhall WRP sludge.

Further indication that digestion adversely affects the dewaterability of sludge was encountered at the JWPCP. Thermal runs JHT 1 and JHT 2, Table XLVIIIA, were made under identical reactor operating conditions but the digested sludge JHT 1 did not dewater as readily as the undigested waste sludge JHT 2. The digested sludge dewatered to 29 to 30 percent TS and exhibited the same "egg shell" effect previously discussed. The undigested sludge was dewatered to approximately 40 percent TS and was consistent throughout.

With regard to the effect of temperature, sludge heat treated at 380°F JHT 2 and JHT 3 yielded filter press cakes of 34 to 45 percent TS while that heat treated at 400°F JHT 4 and JHT 5 produced filter cakes of 46 to 51 percent TS and were competitive to the results obtained on pressure filtration of LPO conditioned sludge. The cake consistency and discharge characteristics obtained by pressure filtration are included in Table XLVIII A, B.

The only centrifuge data collected on dewatering of thermally conditioned WAS was a series of tests made on heat treated WAS at the Saugus-Newhall WRP. The thermal unit was operated at 400°F, 400 psig, and a reactor detention time of 29.2 minutes. The heatrate, following a short detention time in the decant thickener was fed to the 20" x 62" scroll centrifuge at a rate of 30 gpm with a suspended solids concentration of 1.85%. The centrifuge was operated at its maximum pool depth setting, a bowl speed of 2070 rpm and a relative scroll speed of 19 rpm. Cationic polymer was added in the range of 0 to 11.5 lb/ton for conditioning and the results are shown in Figures 83 and 84. Regardless of polymer dosage, the discharged cakes averaged between 20 and 21% TS, but a minimum of 7.5 lb/ton of polymer was required to obtain a centrate containing less than 1500 mg/l of SS. Of extreme importance is the fact that even though the cakes were in excess of 20% TS they were not conveyable and contained large quantities of free moisture. In fact, from a visual standpoint, the discharge cakes were more fluid than the 6% - 7% TS cakes achieved when thickening waste activated sludge.

No dewatering data were collected on run SNHT 3 because problems developed with the sludge transfer pump, and the decant tank contents had to be drained.

Intermediate Pressure Wet Oxidation

Under the intermediate pressure wet oxidation a feed source of thickened WAS from the Saugus-Newhall WRP was fed to the unit at a rate of 1.5 gpm. The feed to the unit had to be drastically reduced because of the limited capacity of the air compressor which supplied the oxygen necessary for oxidation. To maintain a feed of 5-6 gpm under IPO, the amount of air required for oxidation was approximately 60 cfm which was beyond the capacity of the compressor installed on the mobile facility. Consequently, the reactor detention period of cooking time approximated 97 minutes. The reactor temperature and pressure were held constant at 450°F and 500 psig, respectively.

As shown in Table XLIX, volatile solids destruction equalled 41% with an increase in dissolved solids and a decrease in suspended and total solids as the sludge was subjected to intermediate wet oxidation. Thermal solubilization of organics is again evident by an increase in soluble BOD and soluble COD by a factor of ten. Again, odors were present and easily detectable during this phase of the thermal treatment studies.

At the JWPCP, the IPO process was fed at a sludge flow of 3.0 gpm and an air flow of 17.5 cfm. Table L summarizes the operating parameters maintained and the performance data collected. Total COD reduction was measured at 21 percent while the theoretical reduction was calculated to be 37 percent, based on the pounds of oxygen added per pound of COD introduced into the reactor. Solubilization of organics and solids was again evidenced by an increase in the soluble COD concentration from 1600 mg/l to 11,000 mg/l, and an increase in dissolved solids from 2000 mg/l to 10,000 mg/l. Compared to LPO and heat treatment conditioning, the degree of solubilization for this intermediate pressure wet oxidation run was lower and the fact that the measured COD reduction was only 21 percent adds support to the statement made earlier concerning the unpredictable performance of this thermal conditioning unit.

Dewatering of conditioned sludge via vacuum and pressure filtration followed IPO conditioning at the Saugus-Newhall WRP. The data collected are shown in Tables LI and LII. Comparing the results obtained on pressure filtration with those obtained on pressure filtration of LPO and heat treated waste activated sludge, it becomes evident the IPO conditioning enhances the dewaterability of the sludge. Filter press yields increased from 0.50 to 1.05 lb/hr-ft^2 for IPO conditioning while cake remained relatively unchanged at approximately 43% TS. The cake solids obtained under vacuum filtration of the IPO conditioned sludge approximated those obtained for the LPO and heat treated sludges with solids in the range of 27 - 29% TS, but the filter yields increased significantly after IPO conditioning. Filter yields from 12 to 14 lb/hr-ft^2 were obtainable for LPO and heat treatment conditioning.

Vacuum filtration was the only dewatering method employed at the JWPCP because the IPO mode of thermal conditioning was performed during a period when not enough sludge could be processed to run both the vacuum and pressure filters. The data which were collected are included in Table LIII. The vacuum filter was operated at cycle times of 2 to 7 minutes with applied vacuums of 12 to 17 inches Hg. The lower vacuums maintained for these series of tests can only be attributed to physical characteristics of the conditioned sludge. With the vacuum pump operating at full capacity, a vacuum of 20-23 inches Hg is usually achieved, but depending on how well the sludge is picked up by the drum and the porosity of the sludge mat, the vacuum will vary. Resultant cake solids approximated 30 percent while filter yields varied from 2.9 to 4.8 lb/hr-ft^2. These lower cake solids, when compared to the discharges obtained on LPO and heat treated sludges, are probably attributed to the lower concentration of solids in the feed sludge which is the direct result of conditioning a relatively small quantity of sludge during this experiment. The small volume of sludge applied to the decant tank did not allow for a sufficient sludge blanket to develop and, consequently, maximum compaction of the underlying solids was not achieved.

Treatment of the Thermal Liquor

The previous discussion was concerned with the dewaterability of the heat treated sludge after it had been concentrated in a gravity thickener. While this thickened sludge exhibited excellent dewatering characteristics, consideration must at the same time be given to the heat treatment liquor overflowing from the thickener. This liquor has inordinately high concentrations of COD and BOD. The COD approximated 15,000 mg/l with 95 percent being soluble. The BOD approximated 5300 with 95 percent being soluble. The rather small ratio of BOD to COD is both an indication of the exotic nature of the waste and its low rate of biodegradation, and also the lack of an acclimated seed in the BOD test. The degradability of a thermal liquor with activated sludge has been reported in the literature[9,10] along with the numerous problems[9] that have been associated with its direct recycle to the aeration system. These problems are not unexpected, as the difficulties associated with the recycle of the effluent from sludge processing schemes that are heavily laden with suspended solids fines and/or high organic concentrations have continually plagued the successful operation of biological wastewater treatment plants. Certainly some of these problems can be traced to the lack of consideration of the recycle effects on the treatment plant design. However, such accountability in the case of thermal treatment is weakened by the reported problems in predicting the organic solubilization that will be obtained in a full-scale system. Even if one assumes that good scale up can be obtained, it would seem that aerobic treatment of thermal liquor, whether sidestream or recycle to the main plant, would not be the preferred mode. Not only is power spent to transfer oxygen for waste degradation, but an additional biological sludge is generated. Anaerobic treatment would seem preferable over aerobic treatment as not only can energy be recovered in the form of methane gas but the sludge production is minimized.

Operating Experiences with Thermal Conditioning Systems

The Districts' staff experience with thermal conditioning experiments at the Saugus-Newhall WRP and the JWPCP indicates numerous operational problems associated with this process. The particular problems encountered during the Waste Activated Sludge Processing Studies on the pilot scale thermal unit included scale buildup on the walls of the heat exchanger surfaces, odor generation, and various operational failures.

Scaling of the heat exchanger surfaces necessitated an acid flushing after 119 hours of intermittent operation at the Saugus-Newhall WRP. Operation of the same unit at the JWPCP required an acid flushing for scale removal after 101 hours of intermittent operation. The time required to flush the system approximated 8 hours indicating that for the JWPCP at least 5 percent additional capacity must be provided for a full-scale installation to allow for acid flushing downtime.

As mentioned above, a "burnt plastic" or "burnt coffee" odor persisted in and around the pilot plant during these thermal conditioning studies. Odor generation during the Saugus-Newhall WRP studies was more intense than that encountered during the most recent JWPCP studies because the prior studies did not incorporate the use of odor control equipment. During the previous studies the manufacturer indicated that odor control is not a problem provided adequate odor removal devices are installed to treat the vent gases. When the thermal conditioning unit was operated at the JWPCP it was furnished with odor control equipment which significantly reduced process odors, but it was our finding that the characteristic odors were also incorporated in the processed sludge and liquid side-streams. In fact, upon centrifugation of thermally conditioned sludge at the Saugus-Newhall WRP, intense odors were detectable. The turbulence and mixing action in the centrifuge causes a release of the odorous gases, and at times these odors were extremely noxious.

The specific odor control devices utilized during the JWPCP studies included a wet scrubber operated in series with an activated carbon adsorber. The process flow schematic is present in Figure 85. The decant tank vent gases are first processed through a water scrubber which utilizes 20 to 30 gallons of water per 100 gallons of sludge processed. The gases exiting the scrubber are then processed through an activated carbon adsorber which required regeneration after approximately every 12 hours of operation corresponding to a total processed sludge flow of about 4000 gallons.

The liquid waste streams generated by the two-stage odor system were routinely collected and analyzed for total and soluble COD. The average total and soluble COD concentrations of the scrubber waste stream were respectively measured at 35 and 32 mg/l and represent an insignificant COD load if this stream were to be recycled to the plant head works. The carbon regeneration system produced approximately 5 gallons of condensate per regeneration with an average soluble COD concentration of 600 mg/l. This regeneration waste stream constitutes a recycle soluble COD load of less than one pound per ton of solids processed and may be regarded as insignificant.

Samples of the decant tank vent gases and the scrubber and carbon adsorber effluent gas streams were also collected on two separate occasions and quantitatively analyzed for odor intensity. The first set of samples were collected, while the thermal conditioning unit was operated under the heat treatment mode of operation of digested oxygen sludge (JHT 1). The decant tank vent gases had a very strong odor with an average of 17,400 OU/SCF while the exiting gas streams from the scrubber and adsorber had moderate and slight odors, respectively. The average number of odor units measured in the scrubber and adsorber gases were 37 OU/SCF and 15 OU/SCF. Realizing the large volume of water used in the scrubber (20 to 30 percent of the total processed sludge flow), it is not known at this time if a properly sized scrubber will reduce the odors as effectively as the one evaluated. Additionally, for comparison purposes it should be noted that the average emission recorded at the head works of the JWPCP approximates 200 OU/SCF.

The second set of gas samples were collected during intermediate pressure wet oxidation (JIPO) of thickened oxygen waste activated sludge. The decant tank vent gases were measured at 29,000 OU/SCF and were characterized as very strong. The scrubber and adsorber effluent gases were characterized as having "strong" and "very strong odors", while the respective odors were measured at 10,600 and 3,200 OU/SCF. Keeping in mind that the head works to the JWPCP have measured odors of 200 OU/SCF, it becomes apparent that the oxidation process may pose serious odor problems.

Other operational difficulties encountered during the JWPCP studies included: (1) pressure control valve problems after 66 hours of total processing time, (2) compressor failure after 140 hours of sludge processing time, (3) boiler failure and difficulties in maintaining the desired reactor temperature after 200 hours of operation and, (4) corrosion of air lines. In all fairness to the manufacturers, it must be noted that the pilot unit was used at other installations prior to being set up at the JWPCP and the actual processing times incurred prior to these failures may have been substantially longer than those reported. However, the unit was refurbished by the manufacturer prior to its installation at the JWPCP.

There is no reason to suspect these problems were unique to the particular system evaluated, or to the facility at which the tests were conducted. A review[12,13] of various thermal treatment installations indicates similar difficulties. At a number of these installations there were excessive operating costs due to corrosion and scaling conditions.

Corrosion of metals is related to the aggressive properties of water which, in turn, are identified with the solubility relations of calcium carbonate. Calcium carbonate is only slightly soluble in water, but in the presence of carbon dioxide it becomes much more soluble through the formation of bicarbonate. Under given pH conditions, there is an equilibrium among calcium salts, carbonate and bicarbonate radicals, and free carbon dioxide. If the water is oversaturated with calcium, it carries an excess of calcium carbonate which tends to form a protective coating on wetted metallic surfaces. On the other hand, if carbon dioxide is in excess, the tendency is in the reverse, allowing potential solution of the metal.

In general, if a water or sludge has a low pH, it would be expected that some corrosion would occur. Thermal oxidation usually results in low pH conditions due to production of carbon dioxide and organic acids. However, other factors are also important in determining the rate of corrosion. Generally, the corrosion rate will be proportional to the conductivity, which is a measure of the total dissolved minerals. Also, certain anions exert considerable influence on the rate of corrosion; and chloride, in particular, is a strong corrosion catalyst. Sulfate is less corrosive, while bicarbonate tends to reduce corrosiveness of the other two by an inhibitory action. Corrosion is basically a materials problem and should be solvable, or at least controllable, although the cost may be high. In some cases, high corrosion rates have necessitated use of titanium heat exchanger elements.

A notable installation of the wet oxidation process was the Chicago Sanitary District installation which was operated from 1962 to 1972 but has now been replaced by a land disposal system. Operational and maintenance problems were encountered with the process at the Chicago installation, including odor production, scaling in the heat exchangers, and problems of maintaining high pressure equipment.

A summary[15] dealing with the problems associated with thermal treatment follows: "Heat treatment processes have been used for several years in Europe and the United States. It is only within the past several years that significant United States operating and cost data on heat treatment processes have become available. Results from the United Kingdom are now in technical journals. Difficulties with plants in the United Kingdom are generally attributed to the problems of maintaining such items of equipment as high pressure pumps, compressors, and high temperature and pressure reaction systems. Plant difficulties in the United Kingdom were in some cases attributed to the installation of systems at older plants. However, some plants that have ceased operation were specifically designed with new liquid treatment facilities which could accommodate the heat treatment system recirculation loads. The principal cause of process cost and effluent quality problems appears to be a much higher degree of sludge solubilization with heat treatment than was predicted. Available information indicates high costs of operation, maintenance, and effluent quality problems are associated with heat treatment systems. Several United States plants have ceased operating heat treatment systems due to those problems, including Golden Coors, Colorado; Santee, California; and Chattanooga, Tennessee."

Certainly, it must be recognized that research and development of thermal conditioning systems is ongoing, and the operational problems experienced to date may be resolved in the future. However, the alleged problems associated with thermal treatment in the past are authentic and do pose serious operational and maintenance concerns.

WASTE ACTIVATED SLUDGE DRYING

The decision to dry sewage sludges beyond the level attained by normal dewatering methods usually includes the assumption that the dried product will be marketed as a soil conditioner. However, lack of a sufficient market should not necessarily preclude the drying process. The possibility exists, especially for difficult to dewater sludges, that drying might be a viable and required step to render sludges suitable for landfill disposal. Drying is a dehydration process and is accomplished by a variety of methods including direct and indirect heat drying, composting, solvent extraction dehydration, and oil-emersion dehydration. The two methods employed for the JWPCP sludge studies were windrow composting and indirect rotary kiln heat-drying. The operating and process parameters encountered with these two drying processes are discussed below.

Composting

The composting process attempts to create a suitable environment for thermophilic facultative aerobic microorganisms, and in so doing, several criteria must be met to insure successful composting. First, organic solids should be well-mixed. Composting of sewage sludge alone requires that the sludge be blended with previously composted material or bulking agents such as sawdust, straw, wood shavings, or wood chips. This blending process should produce a homogenous porous solids structure in the composting material to enhance aeration. Second, aeration must be sufficient to maintain aerobic conditions in the material. Third, proper moisture content must be maintained. Microorganisms require moisture to function and a moisture content between 45 and 65 percent is generally considered desirable in the composting mixture.

In order to obtain a composting mixture with an initial moisture content of 45 to 65 percent, sludge to be composted must be dewatered prior to blending with compost material. At the time of this study, the practice at the JWPCP was to blend dewatered digested primary sludge at a moisture content of approximately 70 percent with compost material at a moisture content of approximately 30 percent to bring the initial moisture content of the mixture to 60-65 percent. Figure 86 shows temperature, moisture content, and volatility against time from initial blending of dewatered digested primary sludge and previously composted material. Following 10 to 15 days of composting, the mixture obtained a moisture content of approximately 30 percent while the volatility leveled off at 40 percent.

Figure 87 shows the same parameters presented in Figure 86, but reflects compost performance on dewatered digested waste activated sludge. The waste activated sludge was generated at the Saugus-Newhall WRP and was dewatered to a moisture content of 88 percent following anaerobic digestion. To obtain an initial compost moisture content of 65 percent, the dewatered waste activated sludge was blended with previously composted material at a moisture content of 49 percent. After 15 days of composting, the moisture content of the mixture approximated 40 percent while the volatility leveled off at 35 percent. Monitoring of the compost performance continued for 35 days beyond the initial mixing and at the termination of the study the mixture had a moisture content of 20 percent.

Additional composting studies included blending dewatered digested primary and dewatered digested waste activated sludge prior to blending with compost material. The dewatered digested primary and waste activated sludges were mixed at a ratio of 50/50 (solids bases) to bring the combined moisture content to 83 percent. This mixture was then blended with compost material at a moisture content of 49 percent to yield an initial moisture of 65 percent in the compost pile. After fifteen days of operation, the recorded moisture content was 20 percent with a volatility of 30 percent.

It should be noted that all of these experiments reflect summertime operation, and it is expected that the composting time during inclement conditions will approximately double those reported. Nonetheless, it has been shown that waste activated sludge can successfully be composted provided it is thoroughly mixed with enough compost material to yield an initial homogenous moisture of 65 percent or lower.

Mechanical Drying

The class of dryers designated as indirect dryers employ indirect contact of sludge with preheated gases. Common types include rotary dryers and rotary vacuum dryers, and these units can be operated as either batch or continuous type processes. In the batch mode, the dryer is charged with material to be dried and then sealed. A vacuum (approximately 26" Hg) is then applied to the internally charged compartment. Recirculating steam (50-100 psig) from a boiler is passed through a jacketed hollow in the outer shell wall and, in some cases, through the internal central portion of an agitator assembly. The agitator assembly rotates at about 4 rpm and consists of spiral blades which turn the charged material, thereby providing frequent contact of all wetted particles with the heated surfaces. Vapor is removed by vacuum pumps and passed through a condenser prior to discharge.

In the continuous mode no vacuum is applied and the drying process takes place at atmospheric pressure. Material is introduced continuously at one end of the dryer and is discharged continuously at the opposite end. All other process functions are the same except for exhaust gas temperatures, which are much higher in the continuous flow process. This is due to the differences in vaporization temperature requirements at atmospheric pressure versus a vacuum of 6 inches Hg.

A pilot scale 18" x 36" rotary drum vacuum dryer was utilized at the JWPCP[16] to dry four different sludge types. As shown in Table LIV, feed material included digested primary sludge, digested and undigested oxygen waste sludge from the JWPCP, and digested air waste activated sludge from the Saugus-

Newhall WRP. In each case, the sludges were dewatered via scroll centrifugation prior to drying and the initial solids content applied to the dryer is included in Table LIV.

For each test run, the dryer was operated in a batch manner at a jacket temperature of 297°F, an applied vacuum of 28" Hg and allowed to run for approximately four hours. Included in Table LIV are the measured heat requirements expressed in BTU's per pound of water vaporized and the surface heat transfer coefficient (h) for each run.

Depending on the initial moisture content, the initial heat transfer coefficient varies and increases with increasing moisture content. As the sludge dries, the transfer coefficient decreases; effecting an increase in the required drying time and an apparent increase in the unit BTU requirement. Additionally, depending on the sludge characteristics, "balling" or the agglomeration of sludge particles into 2" to 4" balls may occur within the dryer, preventing maximum sludge-dryer contact. The "balling" phenomenon was observed for each of the digested waste activated sludges and causes a decrease in the heat transfer coefficient, an increase in the required drying time and an apparent increase in the unit BTU requirement.

Final moisture content following four hours of drying approximated 3 percent for the digested primary sludge and 60 to 65 on the various biological sludges. It should additionally be noted that during the course of these experiments strong odors were commonly encountered and were characteristic of the odors generated during thermal conditioning studies.

SECTION 5

SYSTEMS EVALUATION

Basically, three waste activated sludge handling schemes have been established as viable alternatives. These are:

1. Flotation thickening, followed by anaerobic digestion, mechanical dewatering, and disposal.
2. Scroll centrifuge thickening, followed by anaerobic digestion, mechanical dewatering, and disposal.
3. Flotation thickening, followed by thermal conditioning, mechanical dewatering, and disposal.

Because of the options available for mechanical dewatering and sludge disposal, the actual number of schemes set up for analysis was twenty. These are presented in Figures 89 through 95.

PROCESS CONSIDERATIONS

Selection of the twenty possible alternatives for subsequent cost effectiveness were based on pilot and full scale data collected by the LACSD and engineering judgment concerning the workability of certain processes. Basket centrifugation was not considered to be a workable thickening process because of the logistics involved when the composite cake is thickened to concentrations of approximately 6 percent TS. At this concentration, approximately half of the basket contents can be skimmed out while the remaining solids have to be plowed out. The skimmed and plowed solids would then have to be blended and mixed prior to pumping to a digester. These procedures coupled with the short run times (13 minutes) encountered with basket centrifuge thickening would present difficult control problems.

Scroll centrifuge thickening was not considered prior to thermal treatment because in our judgment the resultant cake solids of 6 percent TS would be too viscous for optimum thermal efficiency in the heat exchangers and reactor. Such problems have been encountered at Fort Lauderdale, Florida[17] where a heat treatment system was incorporated to condition a disc centrifuge thickened waste activated sludge. At this installation, the centrifuge solids approximated 6 percent TS, and the desired reactor temperatures could not be maintained because of the poor thermal transfer characteristics of the thickened waste activated sludge. Doubling of the heat exchanger capacity was proposed to alleviate the problem, but it remains an untried solution.

A most recent installation of a thermal conditioning unit at Louisville, Kentucky[18] incorporates the use of dissolved air flotation for thickening waste activated sludge prior to treatment. The fact that centrifuges were not employed for thickening adds support to our judgment that concerns exist regarding the mixing and thermal transfer characteristics of the thickened sludge.

Vacuum filtration of digested oxygen waste activated was not considered because of the unsuccessful operation and performance encountered during pilot scale investigations. Pressure filtration, and basket and scroll centrifugation were successful in pilot and/or full scale investigations and were considered workable alternatives for dewatering digested oxygen waste activated sludge.

The disposal methods analyzed included direct landfilling of the dewatered sludge, composting followed by landfilling, and composting followed by disposal to a fertilizer manufacturer. Although the treatment train economics involving pressure filtration, followed by composting and landfilling or delivery to a fertilizer manufacturer were analyzed, a number of uncertainties surround this scheme. Successful pressure filtration requires approximately 800 lb/ton of lime, and it is not known if the dewatered solids are amendable to composting because of the high pH (11.5) obtained. Even if the pressure filter solids can successfully be composted, it is not known if the compost product will be acceptable to a fertilizer manufacturer because of the large amounts of lime included in the product. Utilization of digested centrifuged and composted waste activated sludge by a fertilizer manufacturer was assumed to be a viable ultimate disposal alternative. An analysis of the heavy metals concentration of such a product has indicated that the constituents are equivalent to, if not somewhat lower in concentration than, that presently encountered in the existing primary digested sludge. However, it has been found that the composted waste activated sludge lacks the fiber or bulk characteristics of the composted primary sludge and basically has much finer particle sizes. The finer particle sizes of the waste activated sludge pose a question as to its use as a fertilizer base or soil amendment. Realizing this possible limitation, it was still assumed that the product could be used by a fertilizer manufacturer.

Vacuum and pressure filtration of thermally conditioned sludge were considered for dewatering. Centrifugation was not considered. Thermally conditioned waste activated sludge was dewatered via a 20" x 62" scroll centrifuge, and although the discharge solids approximated 22 percent TS, the nature of these solids were such that they were unconveyable and would pose serious handling problems. If the thermally-conditioned sludge can be centrifuged to yield conveyable discharge solids, the odor problems encountered at the Saugus-Newhall WRP would pose serious environmental concerns. Upon centrifugation of thermally conditioned waste activated sludge, the characteristic thermal odors were intensified by virtue of turbulence in the centrifuge, and at times were extremely noxious. The only disposal method considered was sanitary landfilling of the dewatered thermally treated sludge. Neither composting nor delivery to a fertilizer manufacturer were considered viable alternatives at this time. The composting of thermally conditioned dewatered sludge has not been attempted, but it is probable that odors can emanate from the processed solids when it is stirred or turned in a fashion simulating the action of a mechanical composter. These odors may pose serious environmental concerns.

At the writing of this report significant work had not been conducted regarding the suitability of the thermally treated sludge as a product for use by a fertilizer manufacturer. It has been reported that a thermal sludge is sterile, but concerns are expressed regarding the apparently very fine particle sizes of the sludge, its odor characteristics, and its unknown heavy metal concentrations. Thus, for the purposes of this economic analysis it was assumed that the sludge would not be used by a fertilizer manufacturer.

Treatment of the liquid side-streams inherent to thermal conditioning systems will have to be incorporated in the total processing scheme. If these side-streams were to be recycled to the aeration system, an approximate 12 percent increase in the organic load would occur, necessitating a 12 percent increase in the capacity of the cryogenic unit and the oxygen diffusion equipment. The current secondary system under construction will not allow for this increase in capacity and, as such, it was assumed that anaerobic treatment of the liquid streams would be incorporated.

ECONOMIC ANALYSIS

The data used in preparing these waste activated sludge handling costs were derived from several sources. Equipment manufacturers provided estimates of their respective equipment, and, where possible, costs generated by the LACSD were incorporated in this analysis. The purchase prices and construction costs were all standardized to a consumer price index of 170 and an ENR index of 2400, respectively, and all costs reflect the expenditures necessary for a secondary treatment capacity of 200 MGD. The expected quantity of oxygen waste sludge from this system approximates 106 dry tons per day at a total solids concentration of 1.5 percent.

For simplicity, cost components common to all alternatives such as transfer pumps, conveyors, chemical pumps, etc., are not reflected in these estimates, and as such these are relative rather than actual costs. Power costs were based on $0.03 per KWH while operation and maintenance labor were based on $1,050 per man-month.

Chemical costs were based on the following unit prices: (1) polymer at $4,000 per dry ton, (2) lime as CaO at $40 per dry ton, (3) ferric chloride at $200 per dry ton, and (4) diatomaceous earth at $100 per dry ton. Amortization of capital expenditures were based on an interest rate of 7 percent for a period of 15 years for all equipment with the exception of anaerobic treatment facilities and composting and hauling equipment. Anaerobic digesters and filters were amortized over a period of 25 years while composting and sludge hauling equipment were amortized for 5 years.

The daily chemical requirements (dry tons) for each of the twenty schemes are shown in Table LV. The final solids content and daily wet tons of disposal solids, along with the combined suspended solids concentrations of the process side-streams, are also presented in Table LV.

If composting is feasible after anaerobic digestion and mechanical dewatering, the least amount of solids (wet tons/day) for final disposal would be generated by centrifuging of the digested sludge. The large quantities of lime and diatomaceous earth needed for successful pressure filtration would increase the daily volume of sludge for disposal from 110 wet tons/day (for centrifugation) to 200 wet tons/day. If composting is not possible, then thermal treatment, followed by pressure filtration, will yield 230 wet tons/day of sludge for disposal. Digestion followed by pressure filtration will require the daily disposal of 390 wet tons, while scroll centrifugation of the digested oxygen waste activated sludge will produce 520 wet tons/day for disposal.

A complete summary of the relative costs for each of the twenty schemes is presented in Table LVI. Summarized in Table LVI are the capital expenditures, operation and maintenance costs, and the total annual and unit cost, based on 106 influent dry tons of waste activated sludge. A complete breakdown of the costs associated with each of the unit processes is presented in Appendix A. Included in these appended tables are the total capital and annual capital costs, the amortization period and interest rates used, operation and maintenance labor, maintenance materials, and power, water, and chemical costs. Table A-7 includes the cost of anaerobic filters to treat the liquid side-streams associated with thermal treatment. The anaerobic filter was sized for a 2-day hydraulic detention period, and the costs include those for purchase and installation of filter media. Where applicable, credits were given for the production of methane gas and subtracted from the calculated power costs. A breakdown of composting and land disposal costs is not given in the Appendix because unit costs generated by the operations section of the LACSD for these two unit operations were used. Total annual composting costs were based on $1.80 per dry ton composted, with 33 percent of the total annual cost constituting capital expenditure at an interest rate of 7 percent and an amortization period of 5 years. Hauling or landfill disposal costs were based on a unit price of $.08 per wet ton mile and a dumping price of $2.50 per ton for disposal solids in excess of 25 percent TS and $3.50 per ton for sludges with less than 25 percent TS. Twenty-three (23) percent of the total annual hauling and disposal cost was used to determine the capital expenditure at an interest rate of 7 percent and an amortization period of 5 years for equipment.

A summary of the eight most cost-effective systems is presented in Table LVII. The first four alternate schemes involve composting prior to final disposal. If composting is possible and the product is acceptable by a fertilizer manufacturer, the most economical schemes would incorporate basket or scroll centrifugation after digestion and prior to composting. The unit cost for these alternatives would respectively be $95 and $97 per dry ton of solids processed. If disposal to a fertilizer manufacturer is not possible but composting is still possible, the most effective systems would again involve basket and scroll centrifugation of the digested sludge prior to composting and landfill disposal. The unit cost for these alternatives would respectively be $100 and $102 per dry ton of solids processed.

The last four most cost effective schemes are based on the assumption that composting prior to disposal will not be feasible. In this case, thermal treatment followed by pressure or vacuum filtration and landfill disposal would result in respective unit costs of $96 and $97 per dry ton of solids processed, including the costs associated with anaerobic filtration of the liquid side-streams. It should be noted that the cost estimates for thermal treatment were generated by manufacturers of thermal equipment. Although the capital, power, and operating costs are thought to be reliable, there are serious questions in the mind of the author as to the reliability of the maintenance labor and maintenance material costs supplied.

If digestion and mechanical dewatering prior to landfill disposal were incorporated, then unit costs of $117 and $123 per dry ton would be incurred, respectively, for scroll and centrifugation and pressure filtration. The sludge train involving basket centrifugation of digested waste activated sludge followed by landfilling was not included in this summary because the solids content of the disposal solids would only approximate 11 percent TS and would require a Class I site for disposal. Although the train involving scroll centrifugation of digested oxygen waste activated and digested primary sludge would produce final sludge solids of only 15 percent TS, it was included in the summary pending a ruling as to whether this final product will be accepted at a Class II site. The nature of these solids at a concentration of 15 percent TS are such that no free moisture exists, and they exhibit plastic characteristics.

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TABLE I. FLOTATION PERFORMANCE AND OXYGEN SYSTEM PARAMETERS FOR THE 14 FT^2 DISSOLVED AIR FLOTATION UNIT

	DISSOLVED AIR FLOTATION*						OXYGEN SYSTEM PARAMETERS+			
	FEED			Polymer#	Float	Recovery	MCRT	Total System	SVI (ml/gm)	Sludge Blanket
Date	GPM	% SS	lb SS/hr-ft^2	lb/ton	% T.S.	%	days	Solids (Pounds)	unstirred	ft.
5-7-76	6	1.34	2.9	3.7	4.1	99+	4.2	4104	90	5.5
5-10-76	6	1.34	2.9	2.5	4.0	99+	4.5	4347	76	7.0
5-17-76	6	1.38	3.0	2.9	4.7	99+	4.6	3728	74	4.5
5-19-76	6	1.40	3.0	2.9	5.2	99+	4.1	3961	72	4.5
5-21-76	6	1.56	3.4	2.6	4.8	99+	4.5	4718	82	6.5
5-25-76	6	0.83	2.0	3.2	4.9	99+	5.7	4020	83	6.0
5-26-76	6	1.40	3.0	2.0	3.9	99+	4.1	4122	88	5.0
5-27-76	6	1.32	2.9	2.0	4.2	99+	3.6	3540	75	4.0
6-2-76	6	1.25	2.7	2.7	4.4	99+	4.3	2639	89	2.0
6-3-76	6	1.35	2.9	2.5	4.3	99+	4.8	2452	63	2.5
6-7-76	6	1.09	2.4	3.1	3.5	99+	5.9	3931	87	6.0
6-8-76	6	1.26	2.7	2.7	3.3	99+	4.6	4084	95	5.0
6-9-76	6	1.32	2.9	2.6	3.5	99+	5.5	4558	78	5.0
6-10-76	6	1.35	2.9	2.5	3.7	99+	6.0	4088	86	4.0
6-11-76	6	1.47	3.2	2.3	3.6	99+	5.7	4259	81	4.5
6-14-76	6	1.33	2.9	2.5	3.8	99+	5.6	3437	72	2.5
6-15-76	6	1.46	3.2	2.3	3.9	99+	7.6	3583	79	3.0
6-16-76	6	1.46	3.2	2.3	4.3	99+	6.7	3758	66	3.0
6-17-76	6	1.62	3.6	2.1	3.4	99+	8.0	3541	69	3.0

* For all test runs
  a. Feed Source was O_2 Waste Sludge
  b. Recycle Rate = 26 gpm
  c. Retention Tank Pressure = 58 psig
  d. A/S = 0.03

+ Fourth Stage Aerator speed at 45 rpm.

# Cationic Polymer

TABLE II. SLUDGE THICKENING SUMMARY - OXYGEN WAS

Unit	Suspended Solids Loading	Polymer (lbs/ton)	Cake (% T.S.)	Effluent (mg/l S.S.)	SS Recovery %
Flotation	3-4 lb/hr-ft^2	2-4	3.5	50	99+
Basket Centrifuge	280 lb/hr	>5	5-8	600-800	95
Scroll Centrifuge*	600 lb/hr	7-11	7-9	600-800	95

* Extrapolated from the Saugus-Newhall WAS data.

+ Gravity thickening has been found to be unsuitable for WAS thickening.

# A disc-nozzle centrifuge was examined, but satisfactory operation could not be obtained.

TABLE III. AEROBIC DIGESTION SUMMARY*

Date	Detention Time (days)	Volatile Solids Loading (lb /ft^3-day)	Air Loading Rate (cfm/ft^3)	Volatile Solids Destruction (%)
November 1974	8	0.081	0.043	+
December 1974	8	0.090	0.060	27
January 1975	8	0.096	0.060	23
February 1975	8	0.089	0.060	24
March 1975	8	0.105	0.060	22
April 1975	13	0.070	0.060	27

*All data collected on 100% waste activated sludge from the Saugus-Newhall WRP.

+Start-up and foaming problems prevented an accurate assessment of digester performance

TABLE IV. AEROBIC DIGESTION SUMMARY# - NOVEMBER 1974.

CONSTITUENT	Influent	Effluent	Operating Parameters	
GPD	1,625	1,625	Digester Volume (ft^3)	1,738
Total Solids (%)	1.33	0.84	Detention Time (days)	8
Volatile Solids (%)	77.8	74.4	Vol. Solids ($\frac{lb\ VSS}{ft^3\text{-}day}$)	0.081
Volatile Solids (lb /day)	140	84.9	Air Rate (cfm)	75
pH	6.9	6.1	Air Loading (cfm/ft^3)	0.043
Alkalinity (mg/1)	373	80.2	Temp (°C)	23.5
Total COD (mg/1)	18,000	7,800	Residual D. O. (mg/1)	0.18
Soluble COD (mg/1)	---	62	D.O. Uptake ($\frac{mg/1}{Hr}$)	43.6
Organic Nitrogen (mg/1)	---	--	MLSS (mg/1)	--
NH_3^-N (mg/1)	35.5	10.5	MLVSS (mg/1)	--
NO_2^-N (mg/1)	0.36	0.15	Volatile Solids Dest %	--
NO_3^-N (mg/1)	0.65	113		
SVI	---	--		

#Saugus-Newhall WAS.

TABLE V. AEROBIC DIGESTION SUMMARY# - DECEMBER 1974)

CONSTITUENT	Influent	Effluent	Operating Parameters	
GPD	1,625	1,625	Digester Volume (ft^3)	1,738
Total Solids (%)	1.49	1.16	Detention Time (days)	8
Volatile Solids (%)	78	69.4	Vol. Solids $(\frac{\text{lb VSS}}{\text{ft}^3\text{-day}})$	0.090
Volatile Solids (lb /day)	157	109	Air Rate (cfm)	105
pH	6.9	5.6	Air Loading (cfm/ft^3)	0.060
Alkalinity (mg/l)	437	79	Temp (^{o}C)	23.1
Total COD (mg/l	18,400	9,400	Residual D. O. (mg/l)	1.05
Soluble COD (mg/l)	---	---	D.O. Uptake $(\frac{\text{mg/l}}{\text{Hr}})$	41.8
Organic Nitrogen (mg/l)	---	---	MLSS (mg/l)	8,847
NH_3^-N (mg/l)	49.5	26.2	MLVSS (mg/l)	--
NO_2^-N (mg/l)	0.32	1.30	Volatile Solids Dest %+	26.8
NO_3^-N (mg/l)	1.2	165		
SVI	---	91		

+Includes net increase or decrease in digester total volatile solids.
#Saugus-Newhall WAS.

TABLE VI. AEROBIC DIGESTION SUMMARY# - JANUARY 1975

CONSTITUENT	Influent	Effluent	Operating Parameters	
GPD	1,625	1,625	Digester Volume (ft^3)	1,738
Total Solids (%)	1.54	1.23	Detention Time (days)	8
Volatile Solids (%)	79.8	74.8	Vol. Solids ($\frac{lb\ \ VSS}{ft^3\text{-}day}$)	0.096
Volatile Solids (lb /day)	167	125	Air Rate (cfm)	105
pH	6.9	5.9	Air Loading (cfm/ft^3)	0.060
Alkalinity (mg/l)	402	71	Temp (°C)	22.5
Total COD (mg/l)	17,000	11,600	Residual D. O. (mg/l)	1.20
Soluble COD (mg/l)	530	300	D.O. Uptake ($\frac{mg/l}{Hr}$)	41.4
Organic Nitrogen (mg/l)	1,780	789	MLSS (mg/l)	10,000
NH_3^-N (mg/l)	18.0	48.0	MLVSS (mg/l)	8,100
NO_2^-N (mg/l)	0.25	1.37	Volatile Solids Dest % +	22.8
NO_3^-N (mg/l)	3.9	170		
SVI	---	92		

+Includes net increase or decrease in digester total volatile solids.
#Saugus-Newhall WAS.

TABLE VII. AEROBIC DIGESTION SUMMARY# - FEBRUARY 1975

CONSTITUENT	Influent	Effluent	Operating Parameters	
GPD	1,625	1,625	Digester Volume (ft^3)	1,738
Total Solids (%)	1.43	1.14	Detention Time (days)	8.0
Volatile Solids (%)	80.0	77.3	Vol. Solids ($\frac{lb.\ VSS}{ft^3\text{-}day}$)	0.089
Volatile Solids (lb /day)	155	119	Air Rate (cfm)	105
pH	7.0	5.7	Air Loading (cfm/ft^3)	0.060
Alkalinity (mg/1)	394	30	Temp (°C)	22.6
Total COD (mg/1)	17,900	12,700	Residual D. O. (mg/1)	3.69
Soluble COD (mg/1)	194	149	D.O. Uptake ($\frac{mg/1}{Hr}$)	18.3
Organic Nitrogen (mg/1)	2,755	735	MLSS (mg/1)	9,553
NH_3^-N (mg/1)	6.2	10.3	MLVSS (mg/1)	8,025
NO_2^-N (mg/1)	0.1	4.2	Volatile Solids Dest %+	24.2
NO_3^-N (mg/1)	3.1	149		
SVI	---	106		

+ Includes net increase or decrease in digester total volatile solids.
# Saugus-Newhall WAS.

TABLE VIII. AEROBIC DIGESTION SUMMARY# - MARCH 1975

CONSTITUENT	Influent	Effluent	Operating Parameters	
GPD	1,625	1,625	Digester Volume (ft^3)	1,738
Total Solids (%)	1.67	1.32	Detention Time (days)	8
Volatile Solids (%)	80.7	77.6	Vol. Solids ($\frac{\text{lbs VSS}}{ft^3\text{-day}}$)	0.105
Volatile Solids (lbs/day)	183	139	Air Rate (cfm)	105
pH	7.0	7.2	Air Loading (cfm/ft^3)	0.060
Alkalinity (mg/l)	401	280	Temp (°C)	22.8
Total COD (mg/l)	*20,400	*15,600	Residual D O. (mg/l)	0.20
Soluble COD (mg/l)	*286	*197	D.O. Uptake ($\frac{mg/l}{Hr}$)	46.3
Organic Nitrogen (mg/l)	*1,311	*921	MLSS (mg/l)	12,311
NH_3^-N (mg/l)	23.8	1.21	MLVSS (mg/l)	10,033
NO_2^-N (mg/l)	0.30	0.04	Volatile Solids Dest %+	21.6
NO_3^-N (mg/l)	1.07	3.90		
SVI	---	78		

*1-7 day composite analyzed at San Jose Creek WRP

+Includes net increase or decrease in digester total volatile solids.

#Saugus-Newhall WAS.

TABLE IX. AEROBIC DIGESTION SUMMARY# - APRIL 1975

CONSTITUENT	Influent	Effluent	Operating Parameters	
GPD	1,026	1,026	Digester Volume (ft^3)	1,738
Total Solids (%)	1.68	1.31	Detention Time (days)	12.7
Volatile Solids (%)	80.8	76.3	Vol. Solids ($\frac{\text{lbs VSS}}{\text{ft}^3\text{-day}}$)	0.070
Volatile Solids (lbs/day)	117	85.5	Air Rate (cfm)	105
pH	6.9	6.3	Air Loading (cfm/ft^3)	0.060
Alkalinity (mg/l)	391	126	Temp (°C)	24.8
Total COD (mg/l)	19,132	13,590	Residual D. O. (mg/l)	3.5
Soluble COD (mg/l)	471	248	D.O. Uptake ($\frac{\text{mg/l}}{\text{Hr}}$)	37.5
Organic Nitrogen (mg/l)	1,306	909	MLSS (mg/l)	11,880
NH_3^-N (mg/l)	13.3	3.3	MLVSS (mg/l)	9,646
NO_2^-N (mg/l)	0.08	1.19	Volatile Solids Dest %+	27.4
NO_3^-N (mg/l)	0.40	82.0		
SVI	---	85		

+Includes net increase or decrease in digester total volatile solids.
#Saugus-Newhall WAS.

TABLE X. ANAEROBIC DIGESTION SUMMARY

Location	(% WAS - % primary)	Detention Time (days)	Volatile Solids Loading (lbs/ft^3-day)	Volatile Solids Destruction (%)	Unit Total Gas Production ft^3/lb vs destroyed
Saugus-Newhall	0 - 100	58	0.042	62	15
Saugus-Newhall	23 - 77	32	0.073	53	*
Valencia	23 - 77	43	0.060	50	*
Valencia	31 - 69	20	0.084	54	*
Saugus-Newhall	43 - 57	24	0.085	57	14
Saugus-Newhall	50 - 50	19	0.104	57	*
Saugus-Newhall	64 - 36	28	0.081	54	*
Saugus-Newhall	70 - 30	23	0.120	60	11
Saugus-Newhall	73 - 27	23	0.105	59	*
Saugus-Newhall	100 - 0	46	0.060	51	11
Saugus-Newhall	100 - 0	21	0.125	45	*

*Gas data scarce and questionable.

TABLE XI. MESOPHILIC DIGESTION OPERATING PARAMETERS# (NOVEMBER 1975 - JANUARY 1976)

PARAMETER	Mean	Range
Digester Volume............................ft^3	1600	-----
Feed Rate..................................GPD	551	529-572
Detention Time.............................Days	21.9	21-22.7
Volatile Solids Loading..............lb/ft^3 dy	.085	.08-.091
Volatile Solids Destruction..............lb/dy	42.3	41.6-43
Volatile Solids Destruction..................%	31.6	29.0-34.2
Gas Production.........................ft^3/dy	625	607-642
Gas Production...............ft^3/lb.TVS. Dest.	14.8	14.1-15.4
Gas Analysis............................% CH_4	60.8	60.5-61.1
Temperature.................................°F	93.0	91.5-94.4

#WAS from 0.5 MGD pilot plants at the JWPCP.

TABLE XII. MESOPHILIC DIGESTION SLUDGE DESCRIPTION# (NOVEMBER 1975 - JANUARY 1976)

CONSTITUENT	INFLUENT		EFFLUENT	
	Mean	Range	Mean	Range
Total Solids...................%	4.07	3.88 - 4.26	3.08	2.93 - 3.22
Volatile Solids............% T.S.	72.5	71.7 - 73.3	64.25	64.1 - 64.4
Volatile Solids............lb/dy	136	126 - 146	90.8	83 - 98.5
Fixed Solids..............% T.S.	27.5	26.7 - 28.3	35.75	35.6 - 35.9
Fixed Solids...............lb/dy	51.8	46 - 57.5	50.6	46 - 55.1
pH.........................Units	7.04	6.78 - 7.3	6.94	6.69 - 7.19
Volatile Acids......mg/l CH_3COOH	--	-----------	*	* *
Alkalinity.............g/l $CaCO_3$	--	-----------	4.0	4.0 - 4.1
Total COD..................g/l O	43.1	39.1 - 47.0	28.9	27.0 - 30.8
Soluble COD................g/l O	0.53	0.36 - 0.69	2.4	23.0 - 25.9

Note: # WAS from 0.5 MGD pilot plants at the JWPCP.
*See text.

TABLE XIII. MESOPHILIC DIGESTION-HEAVY METALS ANALYSIS

	O_2 Waste Sludge*		Digester Effluent+	
CONSTITUENT	Total	Soluble	Total	Soluble
Cadmium.........mg/l	1.28	0.002	2.40	0.003
Total Chromium..mg/l	44.0	0.010	65.0	0.020
Copper..........mg/l	20.40	0.010	29.0	0.040
Lead............mg/l	8.32	0.180	27.0	0.030
Nickel..........mg/l	6.0	0.130	9.5	0.170
Silver..........mg/l	0.610	0.001	1.00	0.002
Zinc............mg/l	67.5	0.045	99.0	0.041
Mercury.........mg/l	0.056	<0.0001	0.134	<0.0001

*Feed to Air Flotation Unit at 1.57% T. S.

+Digested Oxygen Sludge at 3.10% T. S.

TABLE XIV. THERMOPHILIC DIGESTION OPERATING PARAMETERS# (MARCH 1976 - JUNE 1976)

PARAMETER	Mean	Range
Digester Volume..........................ft^3	1600	-----
Feed Rate...............................GPD	563	539 - 585
Detention Time..........................Days	21.4	20.5 - 22.3
Volatile Solids Loading.............lb/ft^{3}dy	0.074	0.064 - 0.091
Volatile Solids Destruction.............lb/dy	44.4	35.6 - 55.9
Volatile Solids Destruction.................%	39.4	34.6 - 43.6
Gas Production.......................ft^3/dy	747	646 - 880
Gas Production...............ft^3/lb TVS Dest.	17.0	15.2 - 19.3
Gas Analysis............................% CH_4	60.9	60.2 - 61.6
Temperature.................................°F	120	119 - 120

Note: # WAS from 0.5 MGD pilot plants at the JWPCP.

TABLE XV. THERMOPHILIC DIGESTION SLUDGE DESCRIPTION# (MARCH 1976 - JUNE 1976)

CONSTITUENT	+INFLUENT		+EFFLUENT	
	Mean	Range	Mean	Range
Total Solids..................%	3.45	3.02 - 4.15	2.47	2.34 - 2.69
Volatile Solids..........% T.S.	73.2	72.8 - 73.6	62.6	61.7 - 63.1
Volatile Solids...........lb/dy	119	103 - 145	72.8	65.6 - 83.0
Fixed Solids.............% T.S.	26.9	26.4 - 27.2	37.5	36.9 - 38.3
Fixed Solids..............lb/dy	43.5	37.6 - 52.0	43.5	39.6 - 48.5
pH.......................Units	6.6	6.5 - 6.8	7.4	7.4 - 7.5
Volatile Acids.....mg/1 CH_3COOH	----	-------	*	* *
Alkalinity............g/1 $CaCO_3$	----	-------	4.4	4.2 - 4.5
Total COD.................g/1 O	37.1	33.1 - 46.1	25.7	24.6 - 27.0
Soluble COD...............g/1 O	0.36	0.20 - 0.49	2.96	2.20 - 3.32
Organic N.................g/1 N	2.22	1.95 - 2.43	1.05	0.77 - 1.76
Ammonia...................g/1 N	0.08	0.07 - 0.09	1.48	1.36 - 1.54

Note:+ Data represents monthly averages, March - June, 1976.
WAS from 0.5 MGD pilot plants at the JWPCP.
* See Text.

TABLE XVI. HEAVY METAL ANALYSIS FOR THERMOPHILIC DIGESTION*

CONSTITUENT	RESEARCH DIGESTER			
	INFLUENT#		EFFLUENT+	
	Total	Soluble	Total	Soluble
Cadmium..........mg/l	4.3	0.002	3.1	0.055
Total Chromium...mg/l	100.3	0.090	34.0	0.130
Copper...........mg/l	48.0	0.030	40.7	0.170
Lead.............mg/l	35.8	0.040	29.8	0.480
Nickel...........mg/l	13.1	0.500	13.0	0.280
Silver...........mg/l	2.1	0.005	1.0	0.071
Zinc.............mg/l	95.9	0.050	97.6	0.180
Mercury..........mg/l	0.18	<0.001	0.18	0.0002
Arsenic..........mg/l	0.36	<0.010	0.4	0.890

*Digester operated at 120°F and a 21 day detention time.

#Digester Feed was thickened WAS from the 0.5 MGD pilot plants at 4.36% T.S.

+Digester Effluent at 2.71% T. S.

TABLE XVII. SPECIFIC FILTRATION RESISTANCE DETERMINATIONS ON MESOPHILICALLY DIGESTED OXYGEN ACTIVATED SLUDGE

Fe Cl_3 Dosage lb/ton	Lime Dosage lb/ton CaO	Specific Filtration Resistance 10^{12} cm/g
0	0	1130
160	600	10
200	760	5.0
240	900	8.0
320	1220	0.87
200	0	160
240	0	110
300	0	26
340	0	13
400	0	34
0	460	800
0	600	80
0	760	30
0	900	19
0	1060	22
0	1220	18
0	1360	13
0	1520	12

TABLE XVIII. SPECIFIC FILTRATION RESISTANCE DETERMINATIONS ON MESOPHILICALLY DIGESTED OXYGEN ACTIVATED SLUDGE

Polymer* Dosage lb/ton	Pressure psig	Specific Filtration Resistance 10^{12} cm/g
10	22.	44
20	225	16
10	100	32
20	100	7.3
10	45	32
20	45	6.6

* Cationic Polymer

TABLE XIX. SUMMARY OF PRESSURE FILTRATION DEWATERING OF MESOPHILIC DIGESTED OXYGEN ACTIVATED SLUDGE AT 225 PSIG OPERATING PRESSURE

Ferric Chloride Dosage lb/ton	Lime Dosage lb/ton CaO	pH	Specific Filtration Resistance 10^{12} cm/g	Cake Discharge Total Solids %	Cake Discharge Sludge Solids %	Filter Yield* lb/ft^2-hr	Run Time hrs.
0	1040	12.1	225	14	9	0.08	3
160	580	----	---	17	11	0.12	3
180	700	----	---	23	12	0.13	2
180	700	11.7	24	22	13	0.16	3
180	860	12.3	9.6	28	19	0.23	3
240	860	12.0	9.2	35	21	0.27	3
280	540	----	34	22	14	0.16	3
300	720	10.9	3.7	35	21	0.25	3
300	940	11.5	3.5	37	22	0.28	3
260	1000	----	---	34	22	0.27	3
400	600	9.7	3.0	27	20	0.35	2
360	760	12.0	1.7	34	23	0.44	2
420	780	----	2.5	37	24	0.31	3
400	900	----	1.2	40	24	0.30	3
400	1040	12.1	1.2	34	21	0.27	3

*Based on sludge solids.

TABLE XX. PRESSURE FILTRATION OF COMBINED DIGESTED SLUDGES*

Fraction of Digested Primary Sludge + %	Ferric Chloride Dosage lb/ton	Lime Dosage lb/ton CaO	pH	Specific Filtration Resistance 10^{12} cm/g	Run Time hrs.	Cake Total Solids %	Cake Sludge Solids# %
30	200	720	11.6	11	3	28	18
70	158	600	11.9	1.1	3	45	26
100	94	360	----	70	2	20	15
100	122	380	9.8	34	3	20	14
100	134	420	9.4	23	3	23	14

* Digested primary sludge and mesophilic digested waste oxygen activated sludge.

+ On a solids basis. Balance is mesophilic digested waste oxygen activated sludge.

# Total Cake Solids less conditioning chemicals.

TABLE XXI. SPECIFIC FILTRATION RESISTANCE DETERMINATIONS ON THERMOPHILICALLY DIGESTED OXYGEN ACTIVATED SLUDGE

Ferric Chloride Dosage lb/ton	Lime Dosage lb/ton CaO	pH	Specific Filtration Resistance 10^{12} cm/g
0	0	7.5	1200
200	620	--	61
200	620	10.0	114
200	780	11.8	7.8
200	940	12.0	7.8
300	620	10.4	22
300	780	11.8	3.7
300	780	11.7	4.8
300	940	--	3.1
400	620	10.8	9.0
400	780	12.0	2.0
400	780	--	4.6
400	940	11.7	1.9
Diatomaceous Earth at 2000 lb/ton		--	98

TABLE XXII. PRESSURE FILTRATION OF THERMOPHILIC DIGESTED OXYGEN ACTIVATED SLUDGE

Run	Ferric Chloride Dosage lb/ton	Lime Dosage lb/ton CaO	pH	Specific Filtration Resistance 10^{12} cm/g	Cake Discharge				Filter Yield# lb/ft^2/hr	
					Total Solids %		Sludge Solids + %			
					25mm * Plates	30mm* Plates	25mm* Plates	30mm* Plates	25mm Plates	30mm Plates
1	280	720	11.7	3.1	41	----	----	----	----	----
2	260	800	11.6	21	34	----	----	----	----	----
3	240	760	----	--	30	23	20	14	0.19	0.18
4	300	760	11.5	24	29	21	19	14	0.16	0.15
5	420	780	11.0	13	31	21	19	16	0.15	0.17
6	180	740	12.1	23	26	21	19	15	0.15	0.16
7	240	840	11.8	24	23	20	19	14	0.14	0.13
8	140	720	12.1	40	16	16	18	16	0.12	0.14
9	240	680	10.8	73	11	15	15	11	0.10	0.10
10	240	760	11.1	--	15	11	14	12	0.10	0.12
11	320**	1240	12.3	3.4	31	22	18	15	0.22	0.26
12	240**	900	----	22	23	19	14	12	0.16	0.17

* Designates the thickness of the cakes produced.

+ Total cake solids less conditioning chemicals.

# Based on cake sludge solids.

** 2¼ hour runs, all other runs were 3 hours.

++ Average feed total solids concentration-2.4%.

TABLE XXIII. DIGESTED COMBINED SLUDGE DEWATERING* - 3' X 1' ROTARY DRUM VACUUM FILTER

Filter Test No.		1	2	3
Cloth Medium		Nylon	Nylon	Nylon
Filter Vacuum ("Hg)		22	22	22
Filter Yield (lb/hr-ft²)		0.36	0.47	0.58
Cycle Time (Min:Sec)		2:52	2:52	2:52
Feed	(% TS)	1.94	1.94	1.94
Filtrate	(% TS)	0.11	0.29	0.17
	(% SS)	----	0.06	0.04
Cake	(% TS)	7.2	8.0	8.7
	Discharge	Poor	Poor	Poor
Polymer	lb/ton)	20	0	0
Lime	lb/ton)	400	300	400
$FeCl_3$	lb/ton)	150	150	150

*70% WAS - 30% primary

TABLE XXIVA MESOPHILICALLY DIGESTED OXYGEN SLUDGE DEWATERING ON THE 3' X 1' VACUUM FILTER

FILTER RUN	1	2	3	4	5	6
Vacuum...........in. Hg	24	23	24	22	24	23
Cycle Time......min:sec	2:01	3:50	5:43	2:01	3:47	5:54
Filter Yield..lb/hr-ft^2	0.93	0.69	0.48	3.21	2.43	1.75
Feed...............% TS	2.89	2.89	2.89	2.83	2.83	2.83
Feed...............% SS	2.75	2.75	2.75	2.69	2.69	2.69
Filtrate...........% TS	2.78	2.55	2.20	0.68	0.28	0.27
Filtrate...........% SS	2.64	2.41	2.06	0.54	0.14	0.13
SS Recovery...........%	6.2	10.9	30.5	79.9	94.8	95.2
Cake...............% TS	10.3	11.9	11.2	12.4	13.7	14.9
Cake..........Discharge	Poor	Poor	Poor	Poor	Poor	Poor
Lime..........(lb/ton)	1165	1165	1165	1221	1221	1221
$FeCL_3$.........(lb/ton)	0	0	0	261	261	261

TABLE XXIVB. MESOPHILICALLY DIGESTED OXYGEN SLUDGE DEWATERING ON THE 3' X 1' VACUUM FILTER

FILTER RUN	7	8	9	10	11	12
Vacuum...........in. Hg	22	23	23	23	24	24
Cycle Time......min:sec	2:01	3:47	5:37	2:02	3:51	5:47
Filter Yield..lb/hr-ft^2	3.36	2.40	2.17	2.36	1.31	0.97
Feed...............% TS	2.79	2.79	2.79	2.74	2.74	2.74
Feed...............% SS	2.66	2.66	2.66	2.63	2.63	2.63
Filtrate...........% TS	0.55	0.26	0.18	0.65	0.60	0.45
Filtrate...........% SS	0.42	0.13	0.05	0.54	0.49	0.34
SS Recovery...........%	86.1	95.1	98.1	79.6	81.4	87.2
Cake...............% TS	12.9	13.8	14.3	12.5	12.8	13.7
Cake..........Discharge	Poor	Poor	Poor	Poor	Poor	Poor
Lime...........(lb/ton)	1221	1221	1221	824	824	824
FeCL$_3$..........(lb/ton)	348	348	348	339	339	339

TABLE XXV A. MESOPHILICALLY DIGESTED OXYGEN PLUS DIGESTED PRIMARY SLUDGE DEWATERING ON THE 3' X 1' VACUUM FILTER

FILTER RUN	13	14	15	16	17	18
Vacuum......................in. Hg	22	23	23	23	20	22
Cycle Time.................min:sec	2:00	3:55	5:55	2:00	2:05	4:00
Filter Yield............lb/hr-ft2	3.69	2.54	2.46	0.80	4.45	2.40
Feed.........(% Dig Pri-% Dig O_2)	100-0	100-0	100-0	64-36	64-36	64-36
Feed.........................% TS	2.55	2.55	2.55	2.94	2.94	2.94
Feed.........................% SS	2.40	2.40	2.40	2.75	2.75	2.75
Filtrate.....................% TS	0.35	0.33	0.35	0.61	0.44	0.37
Filtrate.....................% SS	0.20	0.18	0.20	0.42	0.25	0.18
SS Recovery.....................%	91.7	92.5	91.7	84.7	90.2	93.3
Cake.........................% TS	24.6	24.9	25.1	9.2	18.3	18.6
Cake....................Discharge	Fair	Fair	Fair	Poor	Poor	Poor
Lime......................lb/ton	750	750	750	700	1050	1050
$FeCL_3$......................lb/ton	246	246	246	215	215	215

TABLE XXV B. MESOPHILICALLY DIGESTED OXYGEN SLUDGE PLUS DIGESTED PRIMARY SLUDGE DEWATERING ON THE 3' X 1' VACUUM FILTER

FILTER RUN	19	20	21	22	23	24
Vacuum.....................in. Hg	22	23	23	23	23	23
Cycle Time.................min:sec	2:00	4:00	6:00	2:05	3:55	5:55
Filter Yield............lb/hr-ft^2	3.63	2.16	1.52	3.22	2.32	1.68
Feed.........(% Dig Pri-% Dig O$_2$)	50-50	50-50	50-50	38-62	38-62	38-62
Feed.........................% TS	3.18	3.18	3.18	3.30	3.30	3.30
Feed.........................% SS	3.00	3.00	3.00	3.10	3.10	3.10
Filtrate.....................% TS	0.23	0.24	0.23	0.27	0.38	0.33
Filtrate.....................% SS	0.05	0.06	0.05	0.07	0.18	0.13
SS Recovery.....................%	98.2	97.9	98.2	97.6	94.3	96.0
Cake.........................% TS	14.8	15.6	15.6	12.6	13.1	13.0
Cake....................Discharge	Poor	Poor	Poor	Poor	Poor	Poor
Lime......................lb/ton	960	960	960	750	750	750
FeCL$_3$....................lb/ton	196	196	196	192	192	192

TABLE XXVI. PRIMARY SLUDGE DIGESTER OPERATING PARAMETERS*

CONSTITUENT	Influent	Effluent	Operating Parameters	
GPD	8,644	8,644	Digester Volumn (ft^3)	68,845
% T. S.	6.30	2.42	Detention Time (days	57.8
% Volatile	74.8	62.8	Vol. Solid Loading $\frac{lb\ \ VSS}{ft^3\text{-days}}$	0.042
lb Vol. Sol./Day	2,858	1,096	Vol. Solid Destruction %	61.7
			Gas$\left(\frac{ft3}{lb\ \ Vol.\ Sol.\ Destroyed}\right)$	15.3
			Volatile Acid (mg/1)	18.3

*Primary sludge digester operting parameters during periods when digested primary sludge was blended with combined digested sludge or digested waste activated sludge.

TABLE XXVII. LOW PRESSURE WET OXIDATION OPERATING SUMMARY

Thermal Run Designation	Feed Sludge		Air Rate (cfm)	Reactor Parameters		
	Type	GPM		Temp (°F)	Press. (psig)	Td(Min)
SNLPO 1	WAS	4.0	12.5	380	400	36.4
SNLPO 2	WAS	4.0	12.5	400	400	36.4
SNLPO 3	WAS	5.0	12.5	380	400	29.2
SNLPO 4	WAS	5.0	12.5	400	400	29.2
SNLPO 5	WAS	5.5	12.5	380	400	26.5
SNLPO 6	WAS	6.0	12.5	380	400	24.3
SNLPO 7	WAS	6.0	12.5	380	400	24.3
SNLPO 8	77% WAS 23% primary	5.0	12.5	390	400	29.2

TABLE XXVIII. LOW PRESSURE WET OXIDATION DATA SUMMARY

Thermal Run	Feed Solids					Heatrate Solids				Overflow Solids			
	Total		Suspended		Dissolved	Total		Suspended	Dissolved	Total		Suspended	Dissolved
	%	% Vol.	%	% Vol.	%	%	% Vol.	%	%	%	% Vol.	%	%
SNLPO 1	2.16	81.2	2.08	----	0.080	1.62	78.2	0.70	0.92	0.97	84.5	0.038	0.93
SNLPO 2	3.28	81.8	3.20	----	0.085	2.00	81.6	0.68	1.32	1.29	88.7	0.024	1.27
SNLPO 3	2.91	82.2	2.83	84.0	0.085	1.43	85.3	0.25	1.18	1.22	----	0.033	1.19
SNLPO 4	2.53	82.7	2.45	84.0	0.080	1.36	82.2	0.38	0.98	1.02	87.8	0.025	1.00
SNLPO 5	2.62	----	2.48	----	0.140	1.83	----	0.66	1.17	1.21	----	0.078	1.13
SNLPO 6	2.68	82.7	2.60	84.8	0.080	1.99	82.0	0.96	1.03	1.14	87.2	0.059	1.08
SNLPO 7	2.49	81.5	2.34	84.0	0.150	1.87	80.2	0.78	1.09	1.10	84.0	0.046	1.05
SNLPO 8	3.06	81.6	2.85	82.0	0.210	1.26	88.3	----	----	1.17	89.7	-----	----

TABLE XXIX. LOW PRESSURE WET OXIDATION DATA SUMMARY

Thermal Run	Feed					Heatrate					Overflow			
	COD (mg/1)		BOD (mg/1)			COD (mg/1)		BOD (mg/1)			COD (mg/1)			
	Total	Sol.	Sol.	pH	Alk. (mg/1)	Total	Sol.	Sol.	pH	Alk. (mg/1)	Total	Sol.	pH	Alk. (mg/1
SNLPO 1	26,300	160	138	6.9	---	19,300	10,300	3,130	5.1	---	11,200	10,800	5.0	---
SNLPO 2	39,200	408	---	7.3	550	24,600	16,300	-----	5.5	450	15,300	14,700	5.6	450
SNLPO 3	34,600	176	138	6.8	450	17,300	13,400	4,560	5.2	300	13,800	13,300	5.0	200
SNLPO 4	31,400	273	336	6.8	450	16,800	11,900	4,470	5.3	200	12,400	12,000	5.2	250
SNLPO 5	31,500	275	240	---	---	22,000	12,200	4,560	---	---	14,900	12,100	---	---
SNLPO 6	32,000	363	384	---	---	25,300	15,800	4,680	---	---	15,200	13,600	---	---
SNLPO 7	31,200	272	---	6.8	---	23,000	12,100	4,020	5.3	---	12,800	12,000	5.1	---
SNLPO 8	36,500	2,400	---	6.6	---	14,100	12,200	5,190	5.2	---	13,200	11,800	5.0	---

TABLE XXX. LOW PRESSURE WET OXIDATION OPERATING SUMMARY

Thermal Run Designation	Feed Sludge		Air Rate (cfm)	Reactor Parameters		
	Type	GPM		Temp. (°F)	Press. (psig)	Td* (Min)
JLPO 1	O_2 WAS	5.0	14.5	380	400	29
JLPO 2	O_2 WAS	3.5	14.8	380	400	41
JLPO 3	Dig. O_2 WAS+	5.0	14.6	380	400	29
JLPO 4	O_2 WAS	3.5	15.8	400	400	41
JLPO 5	O_2 WAS	5.0	15.4	400	400	29
JLPO 6	O_2 + Air WAS	3.0	16.0	400	400	48

*Hydraulic detention time.

+Thermophilic digested waste oxygen activated sludge.

TABLE XXXI. LOW PRESSURE WET OXIDATION DATA SUMMARY

Thermal Run	Feed Solids			Heatrate Solids			Overflow Solids		
	Total % (% Vol.)	Suspended % (% Vol.)	Dissolved %	Total % (% Vol.)	Suspended % (% Vol.)	Dissolved %	Total % (% Vol.)	Suspended % (% Vol.)	Dissolved %
JLPO 1	3.91 (74.8)	3.70 (---)	0.21	2.95 (74.3)	1.73 (60.4)	1.22	1.19 (---)	0.021 (---)	1.17
JLPO 2	3.26 (74.9)	2.97 (75.3)	0.29	2.81 (72.5)	1.57 (59.3)	1.24	1.24 (74.7)	0.015 (---)	1.23
JLPO 3	2.33 (62.4)	2.08 (63.3)	0.25	2.14 (62.3)	1.30 (48.1)	0.84	1.39 (66.4)	0.55 (50.1)	0.84
JLPO 4	3.35 (---)	3.17 (73.8)	0.18	2.28 (---)	1.24 (54.7)	1.04	1.06 (---)	0.026 (---)	1.03
JLPO 5	3.63 (72.5)	3.43 (73.5)	0.20	2.95 (69.8)	1.66 (55.6)	1.29	1.25 (---)	0.013 (---)	1.24
JLPO 6	3.31 (75.9)	3.09 (77.5)	0.22	2.87 (70.6)	1.57 (55.3)	1.30	1.13 (---)	0.030 (---)	1.14

TABLE XXXII. LOW PRESSURE WET OXIDATION DATA SUMMARY (CON'T)

Thermal Run	Feed				Heatrate				Overflow		
	COD Total (Sol.) mg/1	BOD$_5$ Total (Sol.) mg/1	pH	Alk. (Vol. Acids) mg/1	COD Total (Sol.) mg/1	BOD$_5$ Total (Sol.) mg/1	pH	Alk (Vol. Acids) mg/1	COD Total (Sol.) mg/1	BOD$_5$ Total (Sol.) mg/1	pH
JLPO 1	48,900 (1,290)	9,800 (540)	6.5	1,100 (740)	38,600 (13,700)	11,900 7,300	5.8	410 (330)	14,000 (13,200)	(5,900) (5,800)	5.7
JLPO 2	38,900 (2,050)	7,700 (800)	6.7	720 (560)	33,800 (14,100)	9,300 (4,500)	5.4	270 (610)	14,200 (12,800)	4,900 (4,900)	5.7
JLPO 3	24,100 (2,690)	3,100 (650)	8.1	4,040 (212)	21,400 (7,560)	4,000 (3,100)	8.0	3,090 (440)	15,400 (6,740)	(3,900) (3,200)	7.7
JLPO 4	39,500 (1,030)	11,200 (570)	5.5	740 (204)	23,000 (12,200)	7,500 (4,700)	5.6	480 (604)	12,500 (11,900)	4,800 (4,300)	6.9
JLPO 5	39,300 (1,650)	11,000 (880)	7.2	790 (500)	37,300 (16,000)	10,000 (5,600)	5.7	530 (703)	15,300 (14,700)	11,100 (5,600)	5.6
JLPO 6	42,600 (1,730)	11,600 (760)	6.6	820 (670)	34,300 14,900	9,000 (5,400)	6.0	630 (870)	14,800 (13,400)	5,400 (4,800)	5.9

TABLE XXXIII. THEORETICAL AND MEASURED COD OXIDATION* FOR LPO CONDITIONING

THERMAL RUN	THEORETICAL OXIDATION+	MEASURED OXIDATION
JLPO 1	11.2	21.1
JLPO 2	20.5	13.1
JLPO 3	22.9	11.2
JLPO 4	21.6	41.8
JLPO 5	14.8	5.1
JLPO 6	23.6	19.5

+Theoretical % Oxidation =

$$\frac{\text{scf/min AIR}}{\text{gpm}} \times \frac{\text{gal}}{\text{lbCOD}} \times \frac{0.075 \text{ lb AIR}}{\text{scf}} \times \frac{0.21 \text{ lb } O_2}{\text{lb Air}} \times 100$$

* Effect of steam dilution not considered. See text.

TABLE XXXIV. COLIFORM REDUCTION DATA FOR LPO CONDITIONING

THERMAL RUN	INFLUENT		HEATRATE	
	TOTAL COLIFORM	FECAL COLIFORM	TOTAL COLIFORM	FECAL COLIFORM
JLPO 2	2.3×10^{8}	2.3×10^{8}	2.4×10^{4}	2.4×10^{4}
JLPO 3	9.3×10^{5}	2.3×10^{5}	1.5×10^{6}	9.3×10^{5}
JLPO 4	4.3×10^{8}	4.3×10^{8}	9.3×10^{5}	9.3×10^{5}
JLPO 5	2.3×10^{8}	2.3×10^{8}	4.3×10^{3}	4.3×10^{3}
JLPO 6	2.3×10^{8}	2.3×10^{8}	2.3×10^{4}	2.3×10^{4}

TABLE XXXV A. DEWATERING LPO CONDITIONED WAS* ON THE 3' X 1' ROTARY DRUM VACUUM FILTER

	JLPO 4				JLPO 5				
THERMAL RUN									
Filter Vacuum......(in. Hg)	20	16	17	17	19	19	19	19	15
Filter Yield ...(lb/hr-ft²)	6.2	3.8	3.5	3.0	6.4	4.2	3.5	6.6	5.7
Cycle Time........(Min:Sec)	2:00	4:30	6:55	8:30	1:55	4:45	6:55	1:55	4:45
Feed...............(% T.S.)	11.7	11.7	11.7	11.7	11.7	11.7	11.7	12.7	12.7
...............(% S.S.)	10.6	10.6	10.6	10.6	10.5	10.5	10.5	11.6	11.6
Filtrate...........(% T.S.)	5.2	1.8	3.7	5.1	4.4	3.5	3.5	1.9	1.8
...........(% S.S.)	4.2	0.80	2.6	4.0	3.2	2.3	2.3	0.79	0.64
S. S. Removal...........(%)	61	92.4	75	62	69	78	78	93.2	94.4
Cake...............(% T.S.)	34	34	34	34	34	34	34	32	34
.............Discharge	Good	Good	Good	Good	Good	Good	Good	Fair	Good

*Sludge generated at the JWPCP.

TABLE XXXV B. DEWATERING LPO CONDITIONED WAS ON THE 3' X 1' ROTARY DRUM VACUUM FILTER (CONT.)

THERMAL RUN	JLPO 6		
Filter Vacuum......(in. Hg)	15	15	15
Filter Yield....(lb/hr-ft^2)	5.6	3.7	3.2
Cycle Time.........(Min:Sec)	2:00	4:45	7:05
Feed................(% T.S.)	8.7	8.7	8.7
................(% S.S.)	7.5	7.5	7.5
Filtrate...........(% T.S.)	2.0	3.2	1.8
...........(% S.S.)	0.83	2.0	0.63
S. S. Removal...........(%)	89	74	91.6
Cake................(% T.S.)	32	31	31
..............Discharge	Good	Good	Good

*Sludge generated at the JWPCP.

TABLE XXXVI. DEWATERING LPO CONDITIONED WAS ON THE 8.4 FT^2 FILTER PRESS*

Thermal Run		SNLPO 3	SN LPO 4	SNLPO 6	SNLPO 7
Cloth Medium		Nylon	Nylon	Nylon	Nylon
Filter Pressure	(psig)	102	104	120	110
Filter Yield	(lb/hr-ft^2)	0.47	0.57	0.56	0.40
Press Time	(min)	90	90	90	90
Feed	(% S.S.)	12.7	8.3	10.6	9.4
Filtrate	(% S.S.)	0.09	0.01	0.02	0.01
S. S. Removal	(%)	99	99	99	99
Cake	(% T.S.)	56	49	35	36
	Consistency	Good	Good	Good	Good
	Discharge	Poor	Poor	Poor	Poor

*No precoat or chemical addition.

TABLE XXXVII A. DEWATERING LPO CONDITIONED WAS ON THE 8.4 FT^2 FILTER PRESS*

THERMAL RUN	JLPO 1		JLPO 2		JLPO 3		
Filter Pressure.......(psig)	100	100	100	100	100	100	100
Filter Yield+....(lb/hr-ft^2)	0.88	0.51	0.91	1.01	0.73	0.63	0.50
Press Time............(Min.)	90	180	90	90	90	90	90
Feed.................(% T.S.)	10.1	9.6	8.5	9.8	10.7	9.9	8.6
.................(% S.S.)	9.0	8.5	7.8	9.1	10.0	9.2	7.9
Filtrate.............(% T.S.)	1.1	0.9	1.0	1.0	1.6	0.6	0.6
.............(% S.S.)	0.04	0.05	0.10	0.10	--	--	--
S. S. Removal............(%)	99.6	99.4	98.7	98.9	--	--	--
Cake.................(T T.S.)	39	46	47	48	30	32	29
.............Consistency	Good	Good	Good	Good	Poor	Poor	Poor
..............Discharge	Poor	Poor	Poor	Poor	Poor	Poor	Poor
Precoat#...................	No	No	No	No	No	No	No

* No chemical addition, sludge generated at the JWPCP.

+ Based on feed solids.

# Celite 545 diatomaceous earth.

TABLE XXXVII B. DEWATERING LPO CONDITIONED WAS ON THE 8.4 FT^2 FILTER PRESS* (CONT.)

Thermal Run	JLPO 4		JLPO 5			JLPO 6	
Filter Pressure.......(psig)	100	100	100	100	100	100	100
Filter Yield+....(lb/hr-ft^2)	1.03	1.01	0.99	0.87	1.00	0.92	0.92
Press Time............(Min.)	90	90	90	90	90	90	90
Feed................(% T.S.)	11.7	11.4	11.7	11.9	11.8	9.9	9.5
................(% S.S.)	11.0	10.7	10.7	10.9	10.7	8.7	8.0
Filtrate............(% T.S.)	0.7	0.8	1.2	1.2	1.2	1.21	1.46
............(% S.S.)	0.08	0.05	0.03	0.02	0.02	0.03	0.02
S. S. Removal............(%)	99.3	99.5	99.8	99.8	99.8	99.7	99.8
Cake................(% T.S.)	48	48	48	48	47	47	51
.............Consistency	Good	Good	Good	Good	Good	Good	Good
..............Discharge	Poor	Poor	Poor	Fair	Poor	Fair	Poor-Fair
Precoat#....................	No	No	No	Yes	No	No	No

* No chemical addition, sludge generated at the JWPCP.

+ Based on feed solids.

# Celite 545 diatomaceous earth.

TABLE XXXVIII. HEAT TREATMENT OPERATING SUMMARY

Thermal Run Designation	Feed Sludge		Reactor Parameters		
	Type	GPM	Temp(°F)	Press.(psig)	Td(Min)
SNHT 1	WAS	5.0	380	400	29.2
SNHT 2	WAS	5.0	400	400	29.2
SNHT 3	WAS	6.2	380	400	23.5

TABLE XXXIX. HEAT TREATMENT OPERATING SUMMARY

THERMAL RUN DESIGNATION	Feed Sludge		Reactor Parameters		
	Type	GPM	Temp °F	Press (psig)	Td (min)
JHT 1	Digested O_2 WAS*	5.0	380	400	29
JHT 2	O_2 WAS	5.0	380	400	29
JHT 3	O_2 WAS	3.5	380	400	41
JHT 4	O_2+Air WAS	5.0	400	400	29
JHT 5	O_2+Air WAS	3.5	400	400	41

*Thermophilic digested waste oxygen activated sludge.

TABLE XL. HEAT TREATMENT DATA SUMMARY

Thermal Run	Feed Solids					Heatrate Solids				Overflow Solids			
	Total		Suspended		Dissolved	Total		Suspended	Dissolved	Total		Suspended	Dissolved
	%	% Vol	%	% Vol	%	%	% Vol	%	%	%	% Vol	%	%
SNHT 1	2.74	81.8	2.65	84.0	0.091	2.04	83.8	0.90	1.14	1.24	86.9	0.24	1.00
SNHT 2	2.91	81.8	2.82	83.0	0.095	1.77	84.9	0.63	1.14	1.31	84.9	0.13	1.18
SNHT 3	2.38	83.0	2.29	85.0	0.089	1.32	88.9	0.23	1.09	1.29	87.3	----	----

TABLE XLI. HEAT TREATMENT DATA SUMMARY

Thermal Run	Feed					Heatrate					Overflow			
	COD (mg/1)		BOD(mg/1)		Alk	COD (mg/1)		BOD (mg/1)		Alk	COD (mg/1)			
	Total	Sol	Sol	pH	(mg/1)	Total	Sol	Sol	pH	(mg/1)	Total	Sol	pH	(mg/1)
SNHT 1	34,200	690	---	6.6	650	26,700	12,500	1370	5.7	400	15,000	11,200	6.1	400
SNHT 2	35,200	200	---	6.9	---	22,100	13,600	3870	5.6	---	15,900	14,100	5.5	---
SNHT 3	29,200	360	420	6.3	---	15,700	11,800	4020	5.8	---	------	11,100	5.8	---

TABLE XLII. HEAT TREATMENT DATA SUMMARY

THERMAL RUN	Feed Solids			Heatrate Solids			Overflow Solids		
	Total % (% Vol.)	Suspended % (% Vol.)	Dissolved %	Total % (% Vol.)	Suspended % (% Vol.)	Dissolved %	Total % (% Vol.)	Suspended % (% Vol.)	Dissolved %
JHT 1	2.14 (61.9)	1.90 (62.9)	0.24	1.85 (60.1)	1.13 (45.3)	0.72	1.02 (-)	0.43 (-)	0.59
JHT 2	3.71 (73.4)	3.51 (73.6)	0.20	3.01 (73.3)	1.69 (58.7)	1.32	1.29 (-)	0.031 (-)	1.26
JHT 3	3.17 (74.2)	2.98 (74.5)	0.19	3.00 (72.4)	1.61 (56.9)	1.39	1.37 (-)	0.023 (-)	1.35
JHT 4	3.15 (73.6)	2.95 (75.0)	0.20	2.51 (71.9)	1.27 (56.8)	1.24	1.16 (86.5)	0.036 (83.3)	1.12
JHT 5	2.98 (69.4)	2.78 (70.6)	0.20	2.27 (66.7)	1.20 (49.3)	1.06	0.95 (-)	0.024 (-)	0.93

TABLE XLIII. HEAT TREATMENT DATA SUMMARY (CONT.)

THERMAL RUN	FEED				HEATRATE				OVERFLOW		
	COD Total (Sol.) mg/1	BOD$_5$ Total (Sol.) mg/1	pH	Alk. mg/1	COD Total (Sol.) mg/1	BOD Total (Sol.) mg/1	pH	Alk. mg/1	COD Total (Sol.) mg/1	BOD$_5$ Total (Sol.) mg/1	pH
JHT 1	21,900 (2450)	1800 (600)	8.1	3750	18,700 (7560)	4000 (3100)	8.0	3090	13,000 (7090)	2800 (2600)	7.9
JHT 2	40,700 (1640)	12,300 (920)	5.6	870	37,000 (15,400)	15,000 (6800)	6.1	470	15,500 (15,000)	7400 (5900)	6.6
JHT 3	38,100 (1870)	12,600 (-)	6.4	790	34,700 (17,000)	15,000 (7600)	5.8	610	17,600 (17,000)	8100 (8100)	5.6
JHT 4	35,500 (1280)	--- (---)	6.6	680	23,000 (14,600)	--- (---)	6.0	770	15,400 (15,000)	--- (---)	6.0
JHT 5	31,000 (1830)	10,300 (980)	6.7	830	26,900 (14,100)	7600 (4500)	6.1	580	13,800 (12,800)	5500 (4800)	6.4

TABLE XLIV. HEAT TREATMENT CONDITIONING COLIFORM REDUCTION DATA

THERMAL RUN	INFLUENT		HEATRATE	
	TOTAL COLIFORM	FECAL COLIFORM	TOTAL COLIFORM	FECAL COLIFORM
JHT 1	1.5×10^6	1.5×10^6	2.3×10^6	2.3×10^6
JHT 2	9.3×10^8	9.3×10^8	4.3×10^7	4.3×10^7
JHT 3	2.3×10^8	2.3×10^8	2.3×10^6	2.3×10^6
JHT 4	4.3×10^8	4.3×10^8	9.3×10^4	9.3×10^4
JHT 5	9.3×10^8	2.3×10^8	4.3×10^5	2.3×10^5

TABLE XLV. DEWATERING HEAT TREATED WAS - 3' X 1' ROTARY DRUM VACUUM FILTER

Thermal Run		SNHT 1	SNHT 2
Cloth Medium		Nylon	Nylon
Filter Vacuum	("Hg)	22	21
Filter Yield	(lb /hr-ft^2)	0.72	1.13
Cycle Time	(Min:Sec)	7:07	7:14
Feed	(% SS)	8.63	12.02
Filtrate	(% SS)	0.25	0.20
SS Removal	(%)	97.1	98.3
Cake	(% T.S.)	29.8	37.9
	Discharge	Poor	Good

TABLE XLVI. DEWATERING HT CONDITIONED WAS* ON THE 3' X 1' ROTARY DRUM VACUUM FILTER

THERMAL RUN	JHT 2			JHT 4			JHT 5		
Filter Vacuum.....(in. Hg)	24	23	23	19	18	16	16	18	21
Filter Yield...(lb/hr-ft²)	4.4	2.7	2.3	5.0	3.3	2.9	2.8	3.8	5.9
Cycle Time.......(min:sec)	2:00	4:45	7:10	2:00	4:40	6:45	6:50	4:25	2:00
Feed+ ....(% T.S.)	11.9	11.9	11.9	11.4	11.4	11.4	7.8	7.8	7.8
..............(% S.S.)	11.0	11.0	11.0	10.3	10.3	10.3	6.8	6.8	6.8
Filtrate..........(% T.S.)	1.6	1.6	1.3	1.8	2.1	2.0	1.1	1.2	1.5
..........(% S.S.)	0.74	0.68	0.45	0.62	0.93	0.84	0.14	0.26	0.59
S.S. Removal...........(%)	93.2	93.8	95.9	93.9	91.0	91.8	98.0	96.2	91.3
Cake..............(% T.S.)	33	34	35	34	34	34	35	36	37
.............Discharge	Fair	Good	Good	Fair	Good	Good	Good	Good	Good

*Sludge generated at the JWPCP.

+Underflow from decant thickener.

TABLE XLVII. DEWATERING HEAT TREATED WAS ON THE 8.4 FT^2 FILTER PRESS*

Thermal Run		SNHT 2
Cloth Medium		Nylon
Filter Pressure	(psig)	100
Filter Yield	(lb/hr-ft^2)	0.27
Press Time	(Min)	240
Feed		11.3
Filtrate	(% SS)	0.08
SS Removal	(%)	99
Cake	(% T.S.)	45
	Consistency	Good
	Discharge	Poor

*No precoat or chemical addition

TABLE XLVIII A. DEWATERING HT CONDITIONED WAS ON THE 8.4 FT^2 FILTER PRESS*

THERMAL RUN	JHT 1		JHT 2			JHT 3
Filter Pressure.......(psig)	100	100	100	100	100	100
Filter Yield+....(lb/hr-ft^2)	0.56	0.62	0.82	0.74	0.75	0.56
Press Time............(Min.)	90	90	90	90	90	90
Feed................(% T.S.)	9.1	9.8	11.1	--	11.1	3.28
................(% S.S.)	8.7	9.2	9.9	9.9	10.0	3.10
Filtrate............(% T.S.)	0.5	0.5	1.2	--	1.2	0.20
............(% S.S.)	0.03	---	0.03	0.03	0.03	0.015
S.S. Removal.............(%)	99.6	---	99.7	99.7	99.7	99.5
Cake................(% T.S.)	30	29	42	41	39	34
.............Consistency	Poor	Poor	Fair	Fair	Fair	Poor
...............Discharge	Poor	Poor	Poor	Poor	Fair	Poor
Precoat#....................	No	No	No	No	Yes	No

* No Chemical Addition, sludge generated at the JWPCP.

+ Based on feed solids.

Celite 545 diatomaceous earth.

TABLE XLVIII B. DEWATERING HT CONDITIONED WAS*-8.4 FT^2 FILTER PRESS (CONT.)

THERMAL RUN	JHT 4			JHT 5	
Filter Pressure.......(psig)	100	100	100	100	100
Filter Yield+....(lb/hr-ft^2)	0.80	0.84	0.79	0.84	0.84
Press Time............(Min.)	90	90	90	90	90
Feed................(% T.S.)	10.0	10.8	11.6	8.9	8.9
................(% S.S.)	9.0	9.8	10.6	8.1	8.1
Filtrate............(% T.S.)	1.03	1.03	1.02	0.84	0.84
............(% S.S.)	0.025	0.028	0.022	0.020	0.019
S.S. Removal.............(%)	99.7	99.7	99.8	99.8	99.8
Cake................(% T.S.)	46	47	46	50	51
.............Consistency	Good	Good	Good	Good	Good
...............Discharge	Fair	Poor	Fair-Good	Fair	Fair
Precoat#....................	No	No	Yes	No	No

*No Chemical Addition, sludge generated at the JWPCP.

+Based on feed solids.

#Celite 545 diatomaceous earth.

TABLE XLIX. INTERMEDIATE PRESSURE WET OXIDATION - OPERATIONAL & PERFORMANCE SUMMARY

Operating Summary

Thermal Run Designation	Feed Source		Air Rate (cfm)	Reactor Parameters		
	Type	GPM		Temp(°F)	Press(psig)	Td(Min)
SNIPO	WAS	1.5	15	450	500	97

Data Summary

Constituent		Feed	Heatrate	Overflow
Solids	(% T.S.)	2.91	1.99	0.80
	(% Vol)	81.8	70.6	----
	(% S.S.)	2.79	1.11	0.05
	(% Vol)	84.0	----	----
	(% D.S.)	0.12	0.88	0.75
Total COD	(mg/l)	27,900	29,500	12,500
Soluble COD	(mg/l)	980	10,700	10,900
Soluble BOD	(mg/l)	340	4,400	------
pH		6.6	7.0	7.2
Alkalinity	(mg/l)	750	1,250	----

TABLE L. OPERATION AND PERFORMANCE SUMMARY FOR INTERMEDIATE PRESSURE WET OXIDATION

OPERATING SUMMARY

Thermal Run Designation	Feed Sludge		Air Rate (cfm)	Reactor Parameters		
	Type	GPM		Temp. (°F)	Press. (psig)	Td (Min)
JIPO	O_2+Air WAS	3.0	17.5	430	460	48

DATA SUMMARY

CONSTITUENT	Feed	Heatrate	Overflow
Total Solids........................%	2.83	1.89	0.95
Total Volatile Solids............% TS	71.1	63.4	----
Suspended Solids....................%	2.63	1.04	0.07
Volatile Suspended Solids........% SS	73.2	85.6	----
Dissolved Solids....................%	0.20	0.85	0.87
COD - Total....................mg/l O	29,800	23,700	----
Soluble..................mg/l O	1,630	10,900	----
BOD - Total....................mg/l O	8,800	6,000	----
Soluble..................mg/l O	580	4,000	----
pH.............................Units	6.8	6.8	7.0
Alkalinity................mg/l $CaCO_3$	870	1,000	----

TABLE LI. DEWATERING IPO CONDITIONED WAS ON THE 3' X 1' ROTARY DRUM VACUUM FILTER

Thermal Run		SNIPO	SNIPO
Cloth Medium		Nylon	Nylon
Filter Vacuum	("Hg)	20	23
Filter Yield	/hr-ft^2)	11.8	14.1
Cycle Time	(Min:Sec)	3:14	2:15
Feed	(% S.S.)	13.9	13.9
Filtrate	(% S.S.)	0.50	0.52
SS Removal	(%)	96.5	96.3
Cake	(% T.S.)	29	27
	Discharge	Good	Good

TABLE LII. DEWATERING IPO CONDITIONED WAS ON THE 8.4 FT^2 FILTER PRESS*

Thermal Run		SNIPO
Cloth Medium		Nylon
Filter Pressure	(psig)	100
Filter Yield	lb/hr-ft^2)	1.05
Press Time	(Min)	90
Feed	(% S.S.)	13.9
Filtrate	(% S.S.)	0.02
SS Removal	(%)	99
Cake	(% T.S.)	43
	Consistency	Good
	Discharge	Poor

*No precoat or chemical addition

TABLE LIII. DEWATERING IPO CONDITIONED WAS* ON THE 3' X 1' ROTARY DRUM VACUUM FILTER

THERMAL RUN	JIPO		
Vacuum...............in. Hg	17	14	12
Filter Yield......lb/hr-ft^2	4.78	3.45	2.89
Cycle Time..........min:sec	2:00	4:48	7:11
Feed.................% T.S.	8.42	8.42	8.42
.................% S.S.	7.64	7.64	7.64
Filtrate.............% T.S.	1.22	1.09	1.17
.............% S.S.	0.44	0.31	0.39
S.S. Recovery.............%	94.3	95.9	94.9
Cake.................% T.S.	31	30	31
..............Discharge	Good	Good	Good

*Sludge generated at the JWPCP.

TABLE LIV. MECHANICAL DRYING WITH THE 1.5' X 3.0' ROTARY DRUM DRYER

Feed Sludge				Product TS %	Water Vaporized lbs	BTU/ lb. Water Vaporized	H BTU/ Hr-Ft^2°F
Type	Source	Total Pounds	% TS				
Digested Primary	JWPCP	180	31.4	97.6	1220	1320	8.8
Digested O_2 WAS	JWPCP	200	14.8	34.9	116.4	1560	7.3
Undigested O_2 WAS	JWPCP	160	10.2	41.7	121.0	1360	8.7
Digested WAS	Saug-Nhl	200	9.3	34.9	146.4	1250	10.0

Notes:

1. Jacket temperature for all runs maintained at 297°F
2. Drying time for all runs approximately 4 hours.
3. Applied vacuum of 28 in. Hg.

TABLE LV. SYSTEMS EVALUATION

Sludge Handling		Chemical (T/dy)				Solids to Disposal		Effluent S.S.+
Scheme	Unit Processes	CaO	FeCL3	D.E.*	Polymer	% T.S.	wet tons/dy	(mg/l)
1A	F-D-P-L	33.20	9.96	6.50	0.16	34	390	50
1B	F-D-F-C-L	33.20	9.96	6.50	0.16	60	200	50
1C	F-D-P-C-M	33.20	9.96	6.50	0.16	60	200	50
2A	F-D-B-L	-----	----	----	0.80	11	630	540
2B	F-D-B-C-L	-----	----	----	0.80	60	110	540
2C	F-D-B-C-M	-----	----	----	0.80	60	110	540
3A	F-D-S-L	-----	----	----	1.09	15	520	540
3B	F-D-S-C-L	-----	----	----	1.09	60	110	540
3C	F-D-S-C-M	-----	----	----	1.09	60	110	540
4A	S-D-P-L	33.20	9.96	6.50	0.53	34	390	500
4B	S-D-P-C-L	33.20	9.96	6.50	0.53	60	200	500
4C	S-D-P-C-M	33.20	9.96	6.50	0.53	60	200	500
5A	S-D-B-L	-----	----	----	1.17	11	630	760
5B	S-D-B-C-L	-----	----	----	1.17	60	110	760
5C	S-D-B-C-M	-----	----	----	1.17	60	110	760
6A	S-D-S-L	-----	----	----	1.46	15	520	760
6B	S-D-S-C-L	-----	----	----	1.46	60	110	760
6C	S-D-S-C-M	-----	----	----	1.46	60	110	760
7	F-T-V-L	-----	----	----	0.16	35	270	910
8	F-T-P-L	-----	----	----	0.16	42	230	100

*Diatomaceous Earth

+SS Concentration of the Combined Process Sidestreams

F - Dissolved Air Flotation	P - Pressure Filtration
B - Basket Centrifugation	V - Vacuum Filtration
S - Scroll Centrifugation	C - Composting
D - Anaerobic Digestion	L - Landfilling
T - Thermal Conditioning	M - Fertilizer Manufacturer

TABLE LVI. ECONOMIC ANALYSIS-COST ESTIMATE SUMMARY FOR WAS SLUDGE HANDLING ALTERNATIVES

Sludge Handling		Capital	O & M	Total Annual	Unit Cost
Scheme	Unit Process	($)	($/yr)	($/yr)	($/Inf. ton)
1A	F-D-P-L	18,650,700	2,914,500	4,741,000	123
1B	F-D-P-C-L	18,701,900	2,722,200	4,561,000	118
1C	F-D-P-C-M	18,564,800	2,425,200	4,231,000	109
2A	F-D-B-L	15,697,000	3,106,100	4,640,900	120
2B	F-D-B-C-L	15,932,000	2,261,400	3,851,900	100
2C	F-D-B-C-M	15,857,200	2,099,900	3,671,900	95
3A	F-D-S-L	14,815,700	3,114,000	4,535,000	117
3B	F-D-S-C-L	14,966,500	2,492,800	3,950,900	102
3C	F-D-S-C-M	14,891,700	2,330,800	3,770,900	97
4A	S-D-P-L	17,639,900	3,417,400	5,229,500	135
4B	S-D-P-C-L	17,691,100	3,225,100	5,049,500	131
4C	S-D-P-C-M	17,554,000	2,928,100	4,719,500	122
5A	S-D-B-L	14,687,000	3,609,000	5,129,000	133
5B	S-D-B-C-L	14,922,000	2,994,200	4,570,800	118
5C	S-D-B-C-M	14,847,200	2,832,200	4,390,800	113
6A	S-D-S-L	13,804,900	3,657,100	5,064,100	131
6B	S-D-S-C-L	13,955,700	3,225,700	4,669,000	121
6C	S-D-S-C-M	13,880,900	3,063,700	4,489,000	116
7	F-T-V-L	17,049,400	1,903,300	3,735,500	97
8	F-T-P-L	18,151,700	1,781,300	3,729,800	96

F - Dissolved Air Flotation
D - Anaerobic Digestion
P - Pressure Filtration
C - Composting
L - Landfill Disposal
M - Fertilizer Manufacturer
B - Basket Centrifugation
S - Scroll Centrifugation
T - Thermal Treatment
V - Vacuum Filtration

TABLE LVII. ECONOMIC ANALYSIS - SUMMARY OF MOST COST EFFECTIVE ALTERNATIVES

Sludge Handling		Capital	O & M	Total Annual	Unit Cost*
Scheme	Unit Process	($)	($/yr)	($/yr)	($/inf ton)
2C	F-D-B-C-M	15,857,200	2,099,900	3,671,900	95
3C	F-D-S-C-M	14,891,700	2,330,800	3,770,900	97
2B	F-D-B-C-L	15,932,000	2,261,400	3,851,900	100
3B	F-D-S-C-L	14,966,500	2,492,800	3,950,900	102
7	F-T-P-L	18,151,700	1,781,300	3,729,800	96
8	F-T-V-L	17,049,400	1,903,300	3,735,500	97
3A	F-D-S-L	14,815,700	3,114,000	4,535,000	117
1A	F-D-P-L	18,650,700	2,914,500	4,741,000	123

*Includes Landfill Disposal Costs of $2.50 per ton for sludges greater than 25% TS and $3.50 per ton for final Disposal Solids with less than 25% TS.

F - Dissolved Air Flotation
B - Basket Centrifugation
S - Scroll Centrifugation
D - Anaerobic Digestion
T - Thermal Treatment

P - Pressure Filtration
V - Vacuum Filtration
C - Composting
L - Landfill Disposal
M - Fertilizer Manufacturer

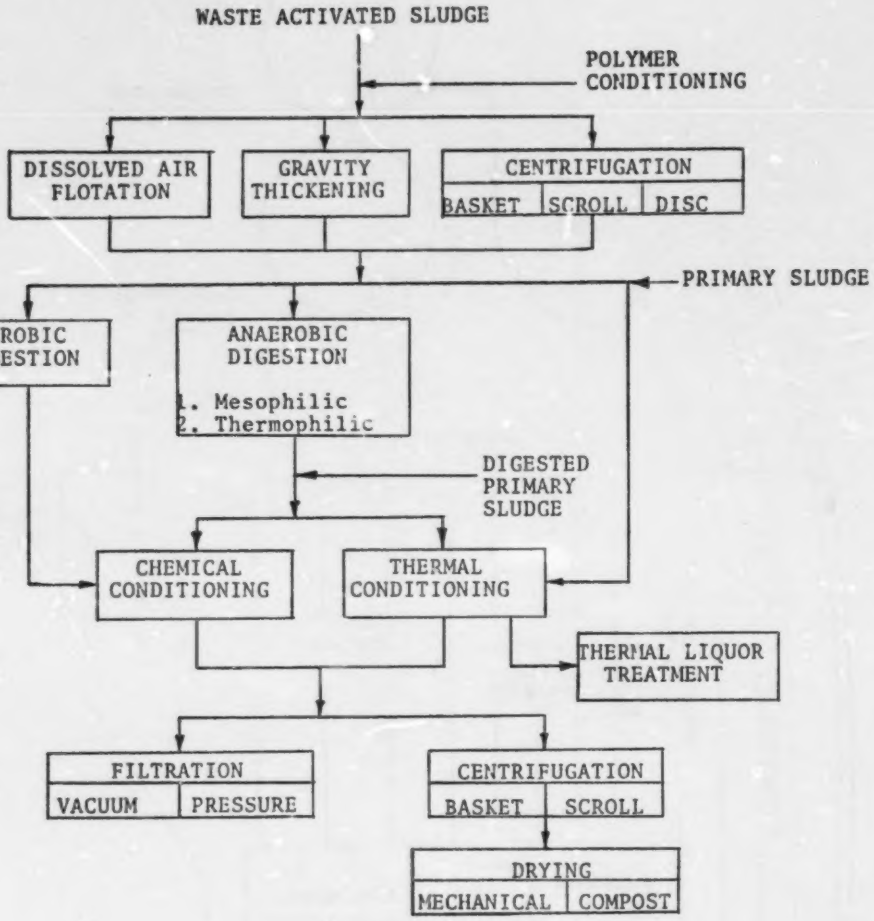

WASTE ACTIVATED SLUDGE
POLYMER CONDITIONING
DISSOLVED AIR FLOTATION
GRAVITY THICKENING
CENTRIFUGATION
BASKET
SCROLL
DISC
PRIMARY SLUDGE
AEROBIC DIGESTION
ANAEROBIC DIGESTION
1. Mesophilic
2. Thermophilic
DIGESTED PRIMARY SLUDGE
CHEMICAL CONDITIONING
THERMAL CONDITIONING
THERMAL LIQUOR TREATMENT
FILTRATION
VACUUM
PRESSURE
CENTRIFUGATION
BASKET
SCROLL
DRYING
MECHANICAL
COMPOST

Figure 1. Research program schematic.

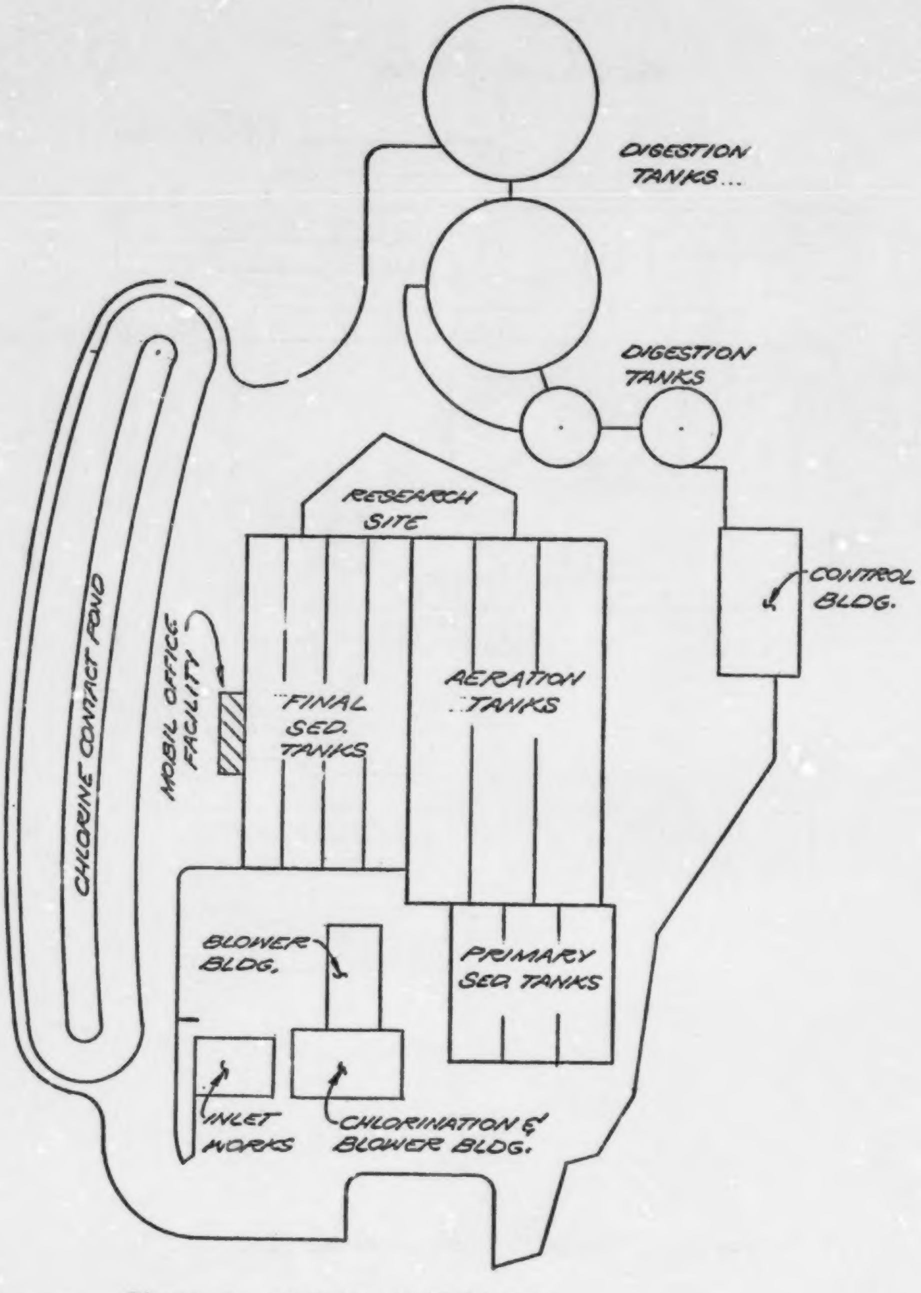

DIGESTION TANKS...
DIGESTION TANKS
RESEARCH SITE
CONTROL BLDG.
CHLORINE CONTACT POND
MOBIL OFFICE FACILITY
FINAL SED. TANKS
AERATION TANKS
BLOWER BLDG.
PRIMARY SED. TANKS
INLET WORKS
CHLORINATION & BLOWER BLDG.

Figure 2. District 26 WRP layout.

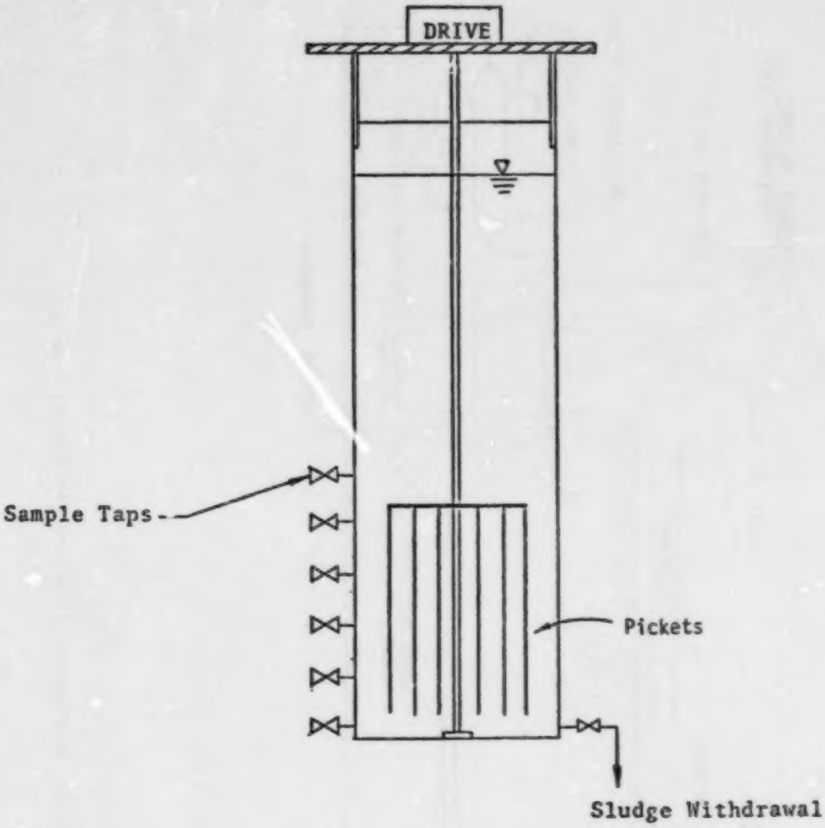

DRIVE
Sample Taps
Pickets
Sludge Withdrawal

Figure 3. 22" diameter x 72" gravity thickener.

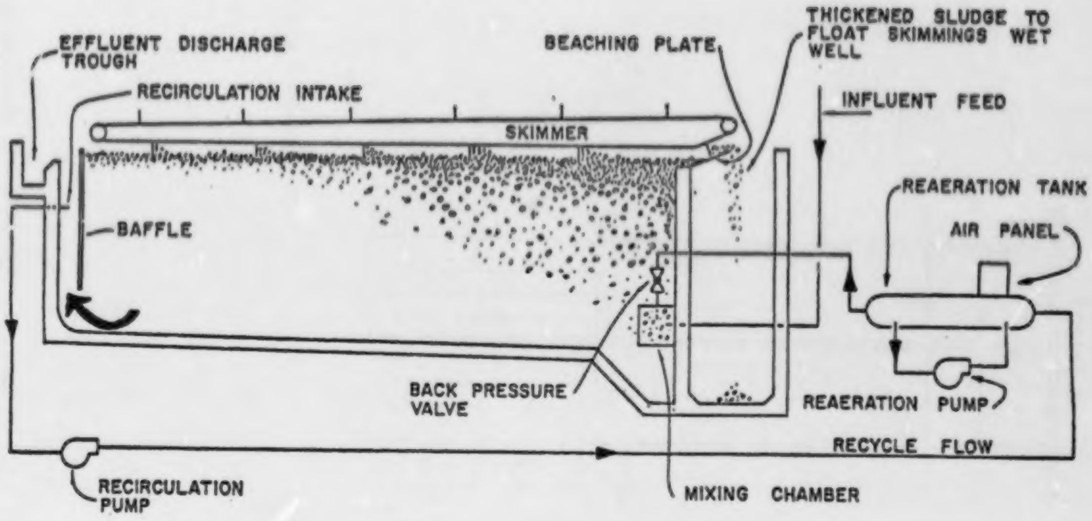
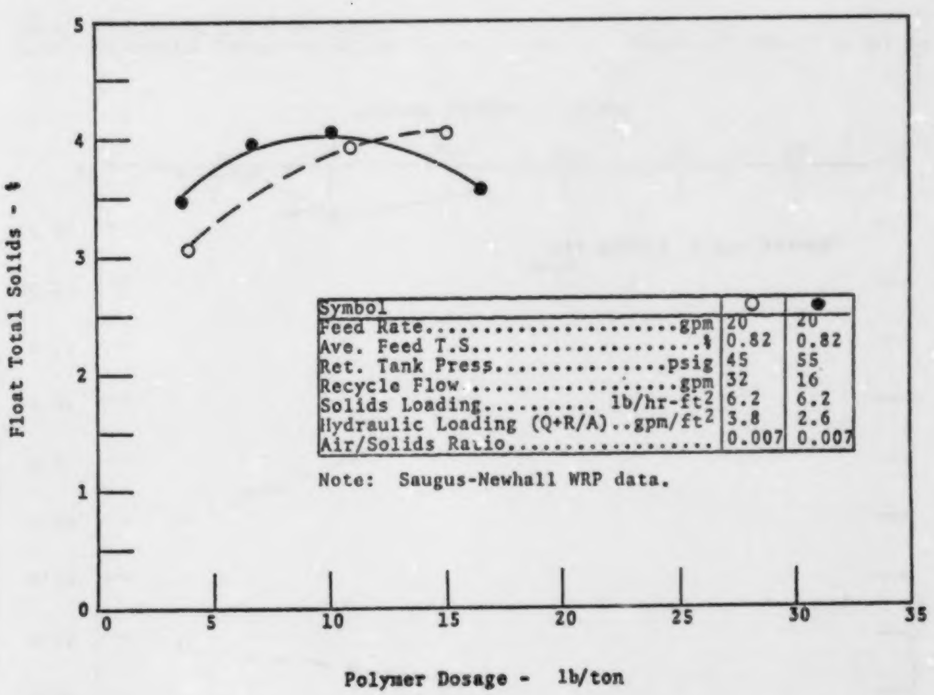

EFFLUENT DISCHARGE TROUGH
RECIRCULATION INTAKE
BEACHING PLATE
THICKENED SLUDGE TO FLOAT SKIMMINGS WET WELL
INFLUENT FEED
SKIMMER
BAFFLE
REAERATION TANK
AIR PANEL
BACK PRESSURE VALVE
REAERATION PUMP
RECYCLE FLOW
RECIRCULATION PUMP
MIXING CHAMBER

Figure 4. Rectangular dissolved air flotation unit.



Symbol	○	●
Feed Rate........................gpm	20	20
Ave. Feed T.S....................%	0.82	0.82
Ret. Tank Press................psig	45	55
Recycle Flow ....gpm	32	16
Solids Loading.......... lb/hr-ft^2	6.2	6.2
Hydraulic Loading (Q+R/A)..gpm/ft^2	3.8	2.6
Air/Solids Ratio..................	0.007	0.007

Note: Saugus-Newhall WRP data.

Figure 5. Float solids vs. polymer dosage for sludge thickening on the 14 ft^2 rectangular dissolved air flotation unit.

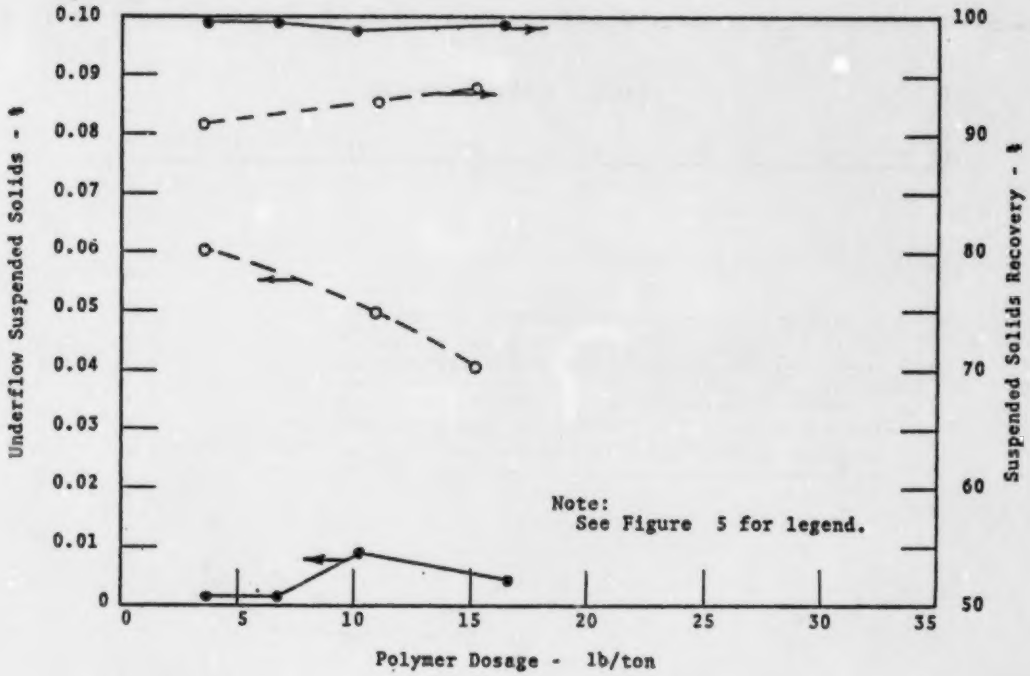
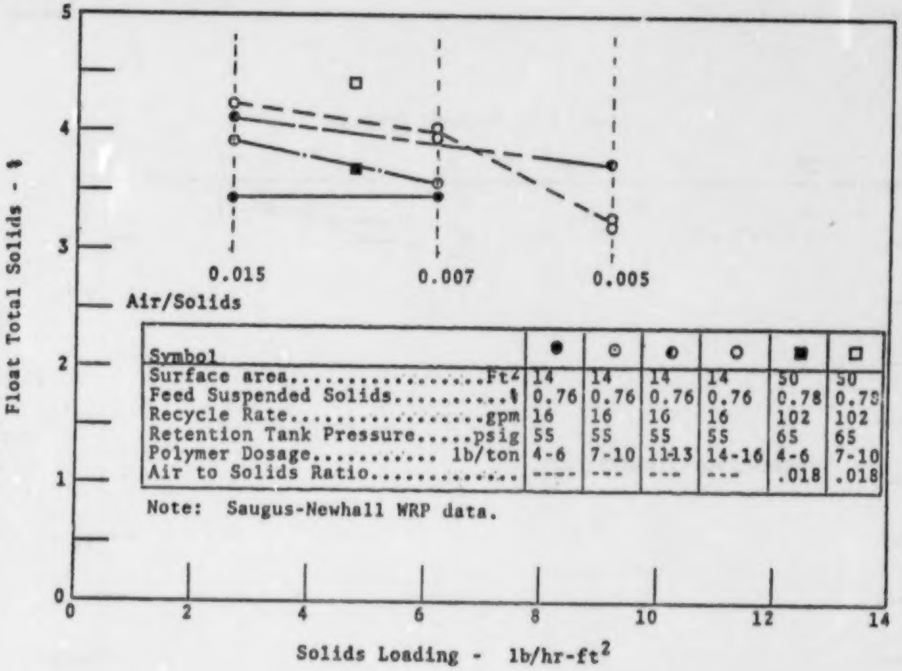


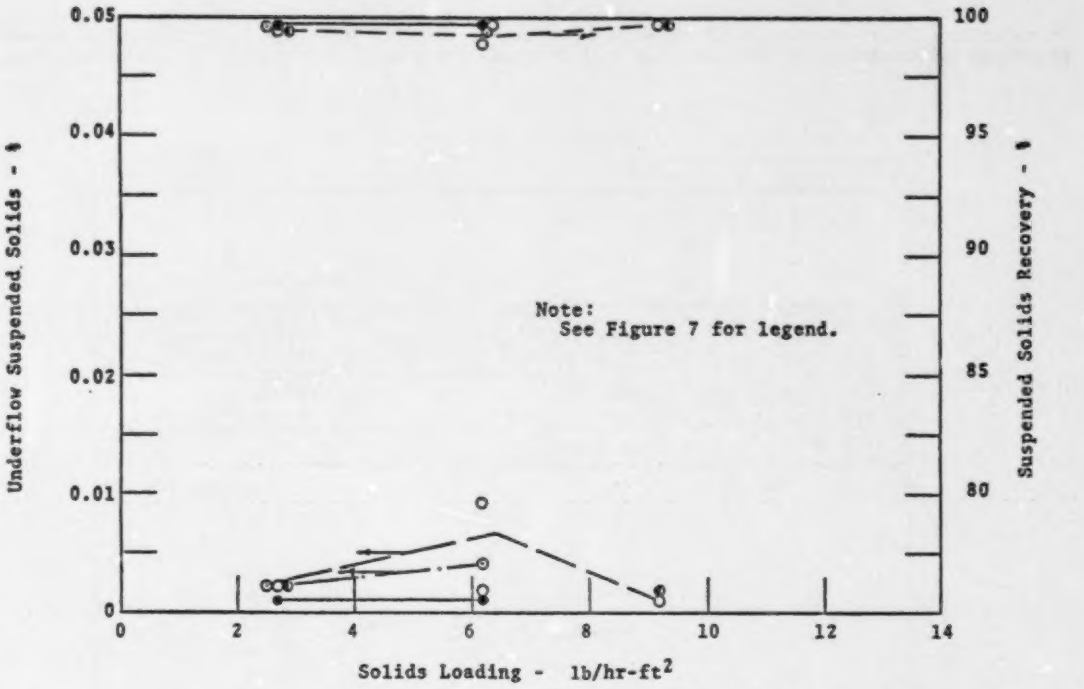
Figure 6. Underflow quality and % removal vs. polymer dosage for sludge thickening on the 14 ft^2 rectangular dissolved air flotation unit.



Symbol	●	⊙	◐	○	■	□
Surface area..................Ft²	14	14	14	14	50	50
Feed Suspended Solids..........%	0.76	0.76	0.76	0.76	0.78	0.78
Recycle Rate..................gpm	16	16	16	16	102	102
Retention Tank Pressure.....psig	55	55	55	55	65	65
Polymer Dosage........... lb/ton	4-6	7-10	11-13	14-16	4-6	7-10
Air to Solids Ratio............	----	---	---	---	.018	.018

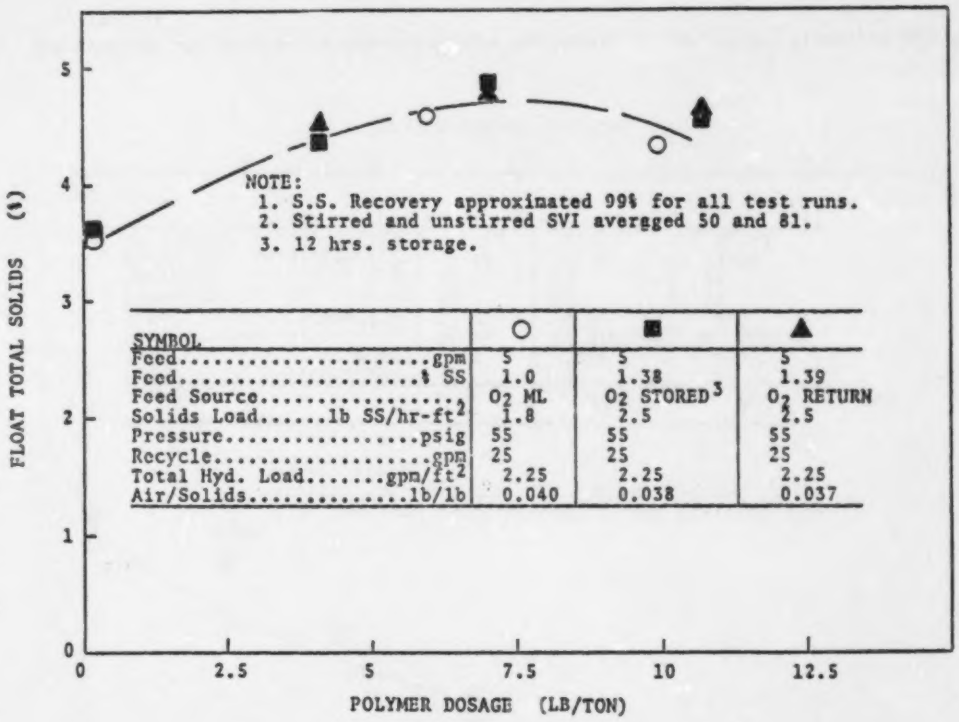
Note: Saugus-Newhall WRP data.

Figure 7. Float solids vs. solids loading for sludge thickening on the rectangular dissolved air flotation units.



Note:
See Figure 7 for legend.

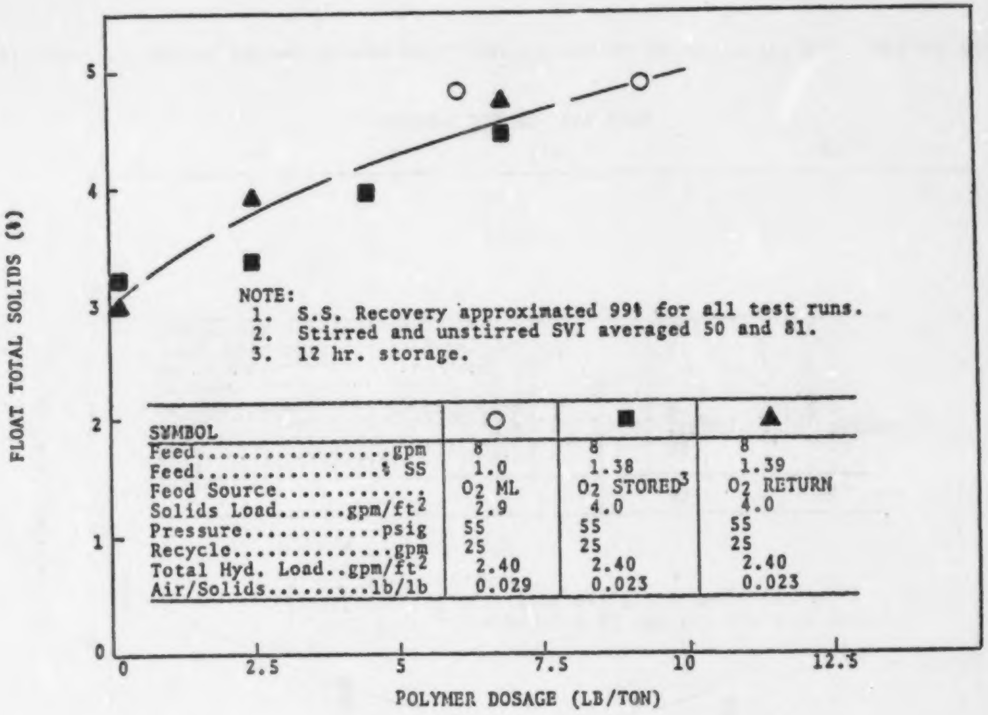
Figure 8. Underflow quality & % removal vs. solids loading for sludge thickening on the rectangular dissolved air flotation units.



NOTE:
1. S.S. Recovery approximated 99% for all test runs.
2. Stirred and unstirred SVI avergged 50 and 81.
3. 12 hrs. storage.

SYMBOL	○	■	▲
Feed....................gpm	5	5	5
Feed....................% SS	1.0	1.38	1.39
Feed Source..................	O_2 ML	O_2 STORED[3]	O_2 RETURN
Solids Load......lb SS/hr-ft^2	1.8	2.5	2.5
Pressure..................psig	55	55	55
Recycle...................gpm	25	25	25
Total Hyd. Load.......gpm/ft^2	2.25	2.25	2.25
Air/Solids..............lb/lb	0.040	0.038	0.037

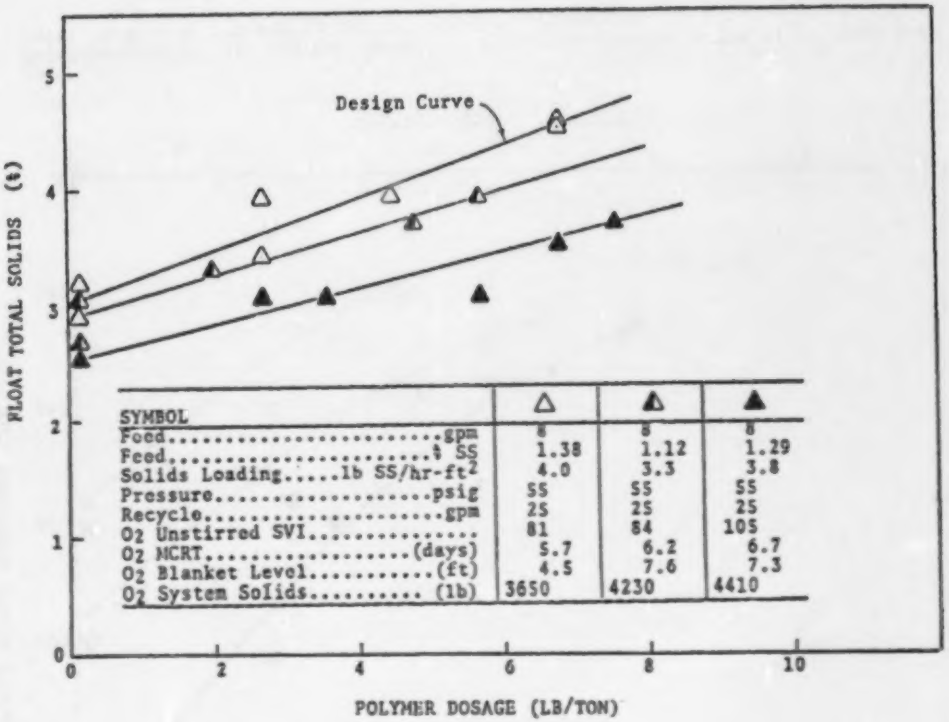
Figure 9. Float solids vs. polymer dosage for sludge thickening on the 14 ft^2 dissolved air flotation unit.



NOTE:
1. S.S. Recovery approximated 99% for all test runs.
2. Stirred and unstirred SVI averaged 50 and 81.
3. 12 hr. storage.

SYMBOL	○	■	▲
Feed..................gpm	8	8	8
Feed...................% SS	1.0	1.38	1.39
Feed Source.............	O_2 ML	O_2 STORED[3]	O_2 RETURN
Solids Load......gpm/ft^2	2.9	4.0	4.0
Pressure............psig	55	55	55
Recycle..............gpm	25	25	25
Total Hyd. Load..gpm/ft^2	2.40	2.40	2.40
Air/Solids.........lb/lb	0.029	0.023	0.023

Figure 10. Float solids vs. polymer dosage for sludge thickening on the 14 ft^2 dissolved air flotation unit.



SYMBOL	△	◭	▲
Feed........................gpm	8	8	8
Feed........................% SS	1.38	1.12	1.29
Solids Loading.....lb SS/hr-ft^2	4.0	3.3	3.8
Pressure...................psig	55	55	55
Recycle......................gpm	25	25	25
O_2 Unstirred SVI...............	81	84	105
O_2 MCRT...................(days)	5.7	6.2	6.7
O_2 Blanket Level.............(ft)	4.5	7.6	7.3
O_2 System Solids........... (lb)	3650	4230	4410

Figure 11. Float solids vs. polymer dosage for sludge thickening on the 14 ft^2 dissolved air flotation unit; flotation and oxygen system operating parameters.

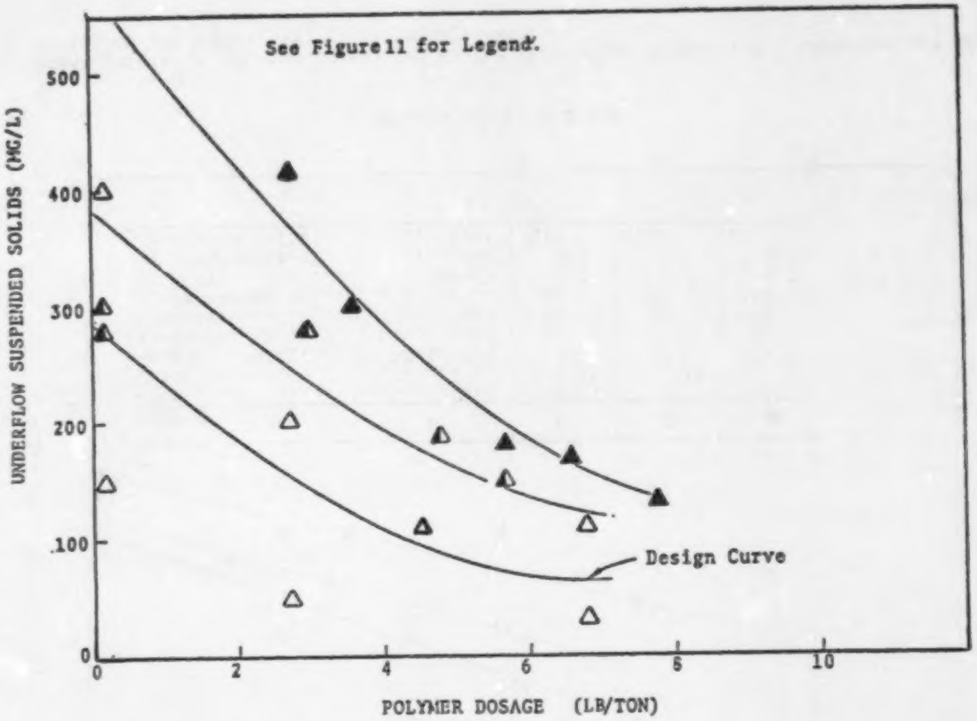


Figure 12. Underflow quality vs. polymer dosage for sludge thickening on the 14 ft^2 dissolved air flotation unit; flotation and oxygen system operating parameters.

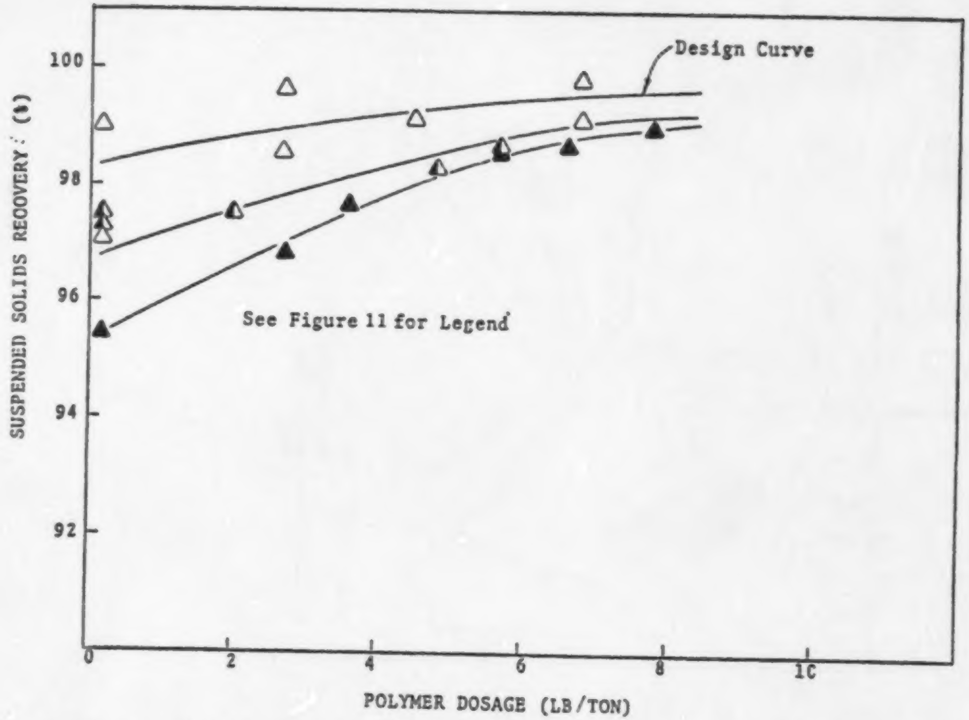


Figure 13. Suspended solids recovery vs. polymer dosage for sludge thickening on the 14 ft^2 dissolved air flotation unit; flotation and oxygen system parameters.

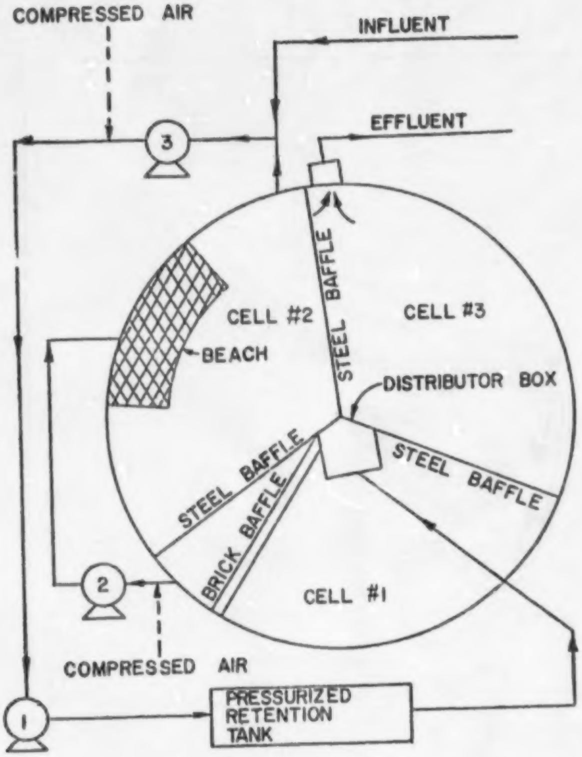

COMPRESSED AIR
INFLUENT
EFFLUENT
3
STEEL BAFFLE
CELL #2
CELL #3
BEACH
DISTRIBUTOR BOX
STEEL BAFFLE
STEEL BAFFLE
BRICK BAFFLE
2
CELL #1
COMPRESSED AIR
1
PRESSURIZED RETENTION TANK

Figure 14. 28 Ft^2 Circular Dissolved Air Flotation Unit.

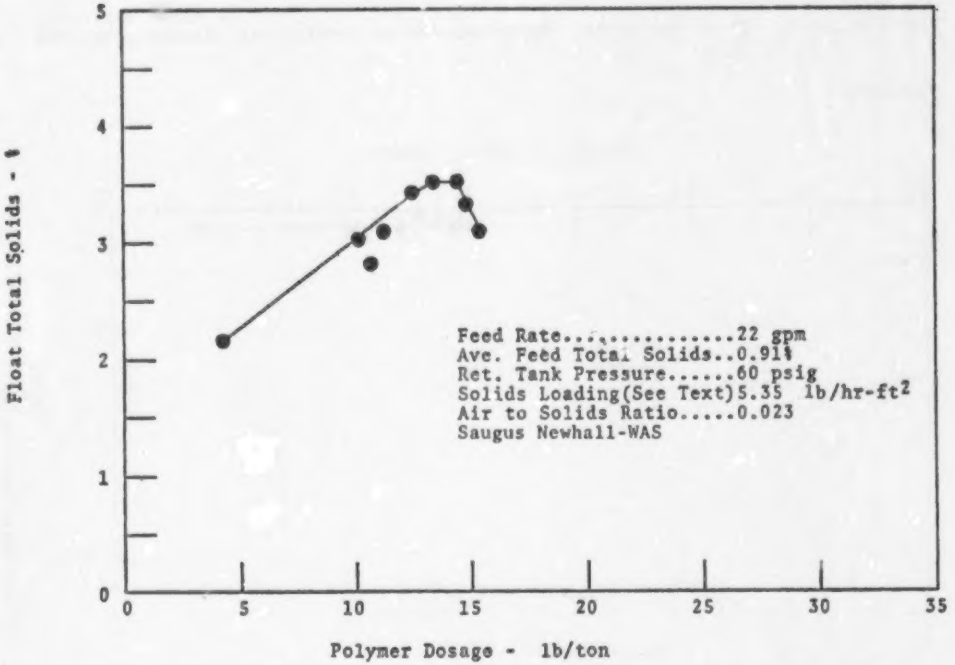
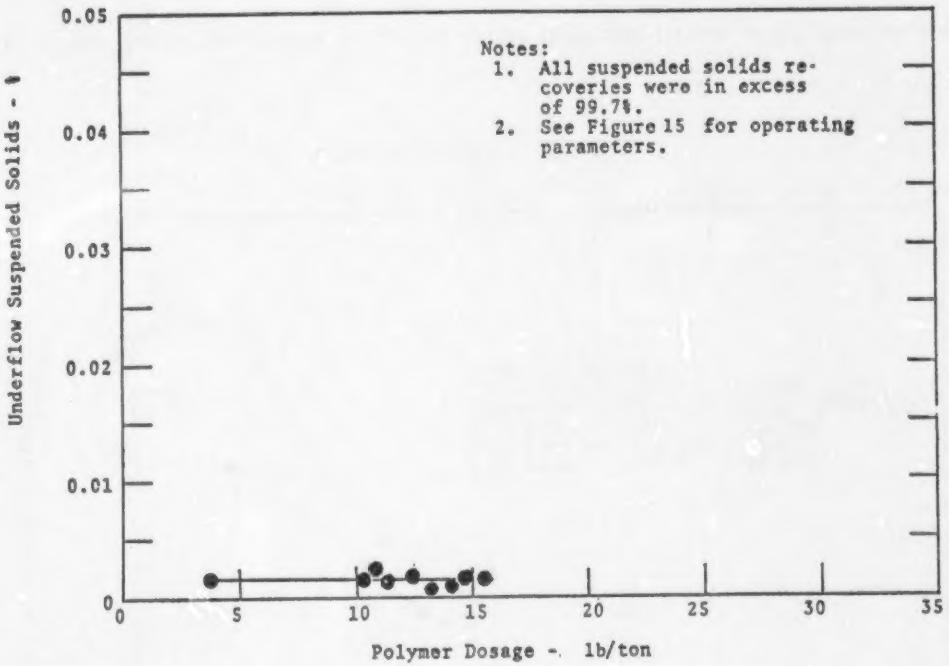


Figure 15. Float solids vs. polymer dosage for sludge thickening on the 28 ft^2 circular dissolved air flotation unit.



Notes:
1. All suspended solids recoveries were in excess of 99.7%.
2. See Figure 15 for operating parameters.

LACSD/76

Figure 16. Underflow quality vs. polymer dosage for sludge thickening on the 28 ft^2 circular dissolved air flotation unit.

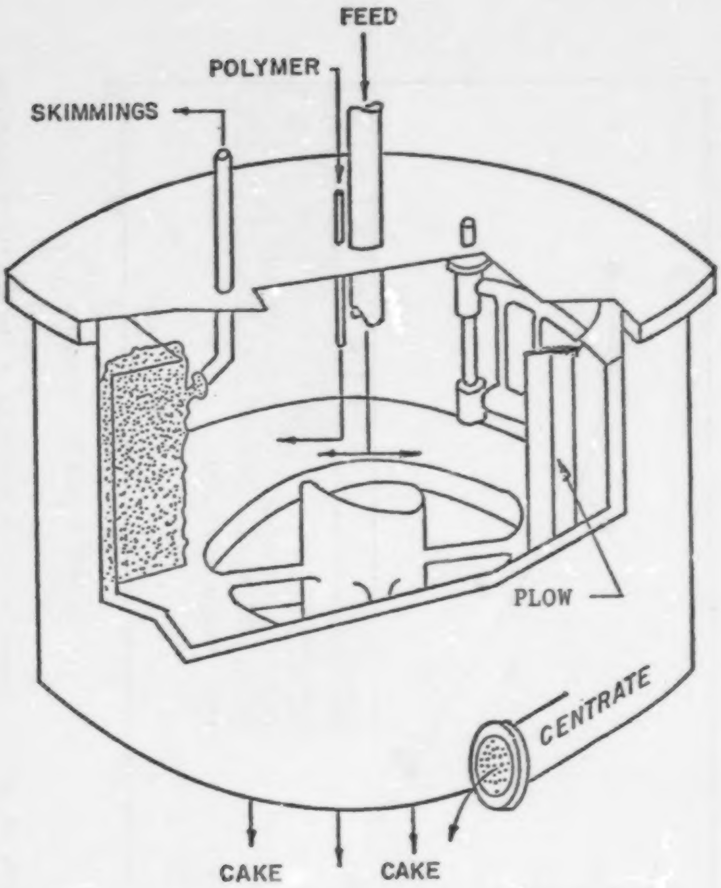
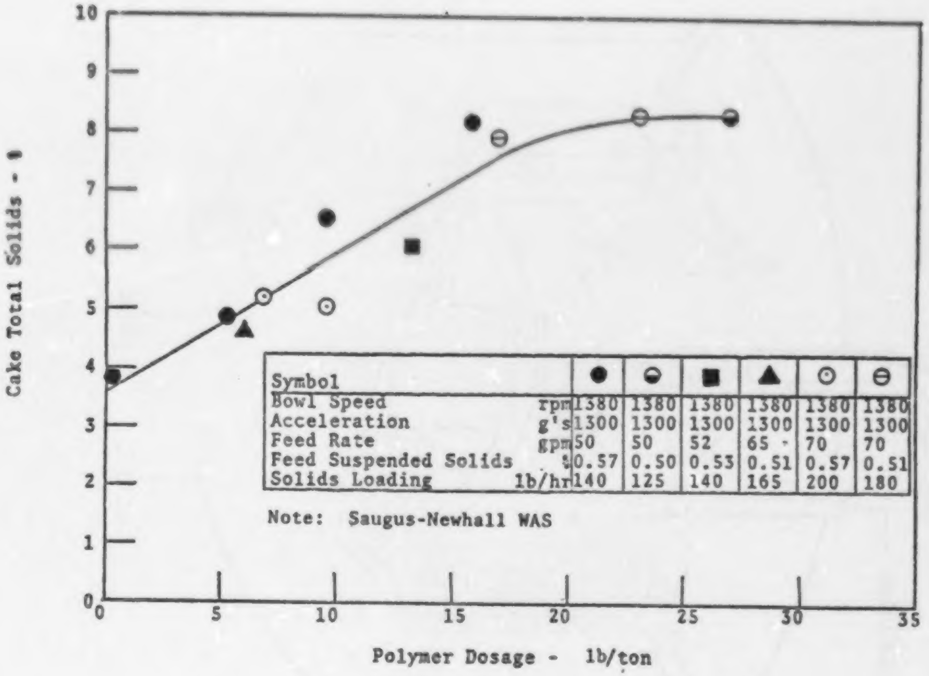

FEED
POLYMER
SKIMMINGS
PLOW
CENTRATE
CAKE
CAKE

Figure 17. Basket centrifuge.



Symbol		●	◒	■	▲	⊙	⊖
Bowl Speed	rpm	1380	1380	1380	1380	1380	1380
Acceleration	g's	1300	1300	1300	1300	1300	1300
Feed Rate	gpm	50	50	52	65 -	70	70
Feed Suspended Solids	%	0.57	0.50	0.53	0.51	0.57	0.51
Solids Loading	lb/hr	140	125	140	165	200	180

Figure 18. Cake solids vs. polymer dosage for sludge thickening on the 48" basket centrifuge.

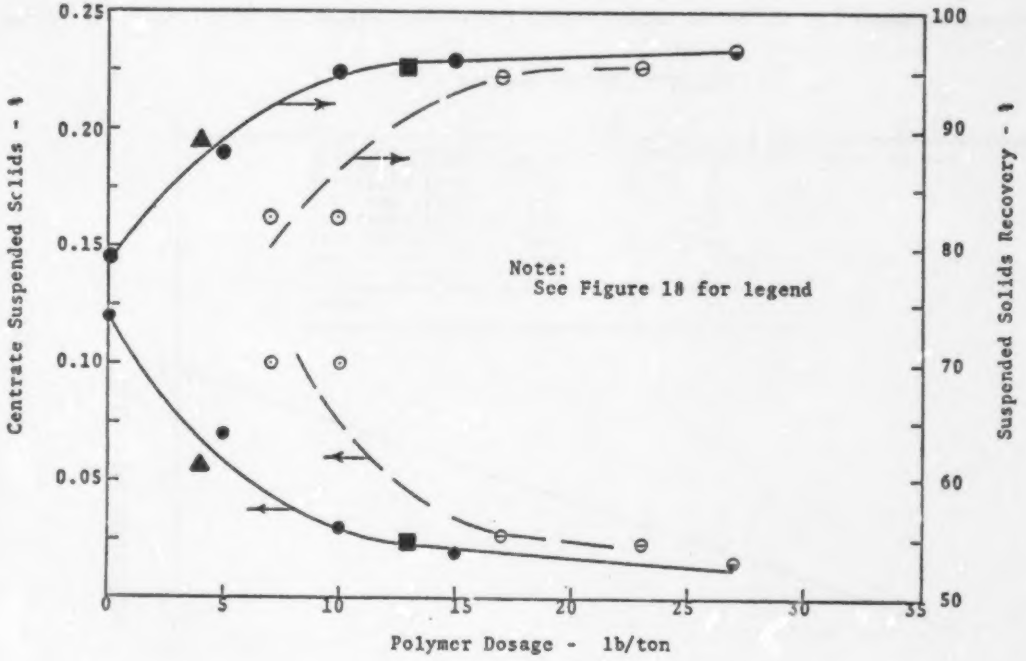


Figure 19. Centrate quality and % removal vs. polymer dosage for sludge thickening on the 48" basket centrifuge.

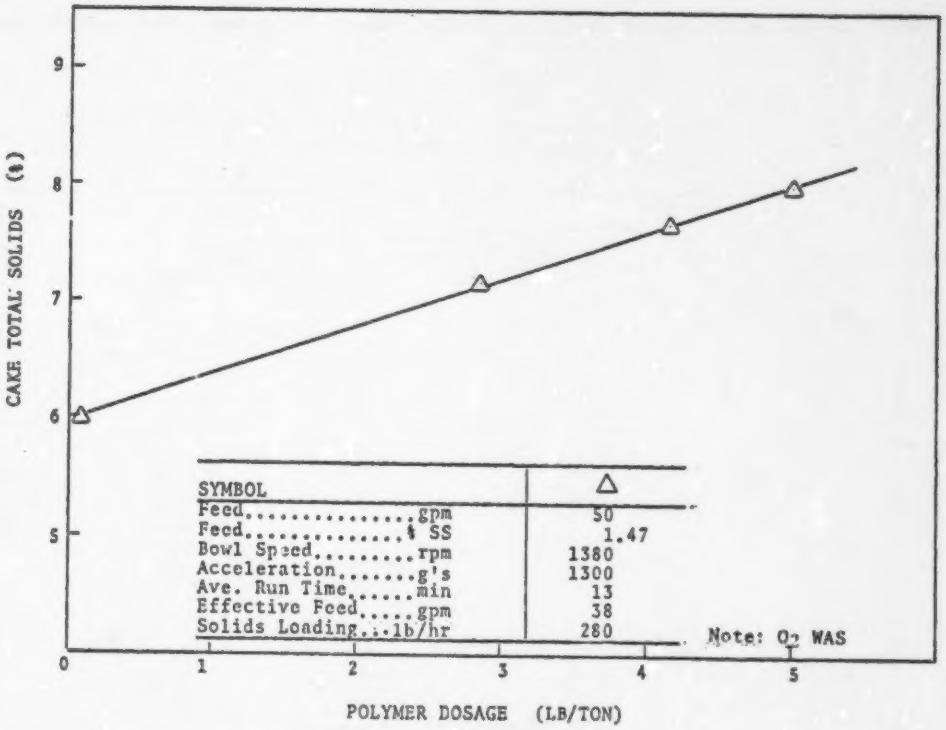


Figure 20. Cake solids vs. polymer dosage for sludge thickening on the 48" basket centrifuge.

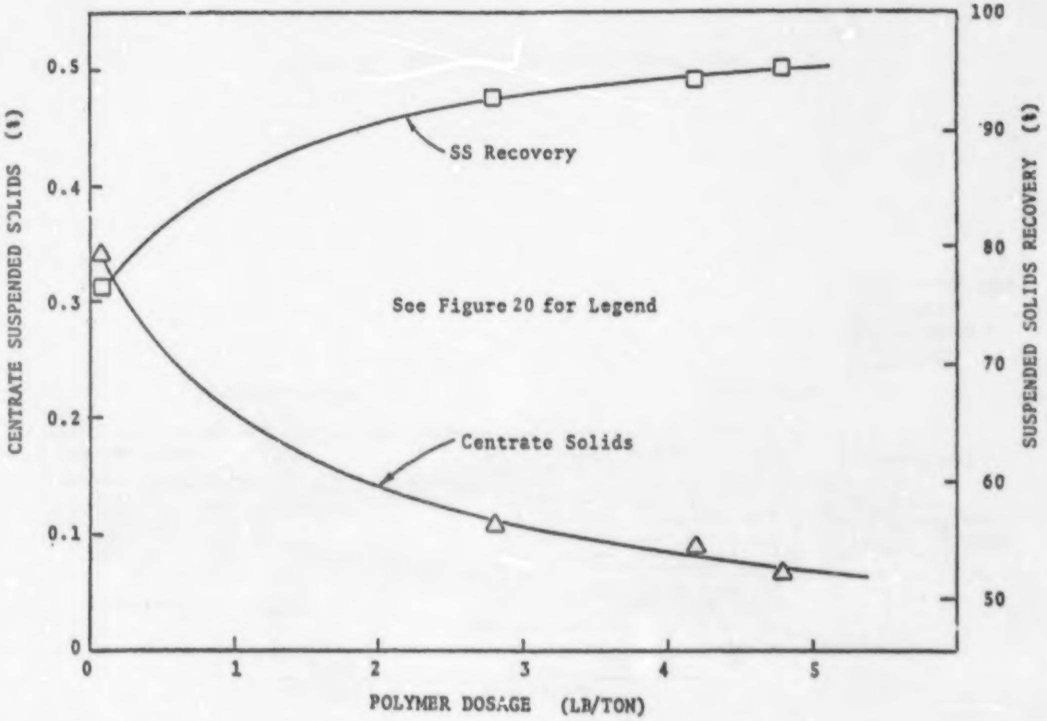


Figure 21. Centrate quality & SS recovery vs. polymer dosage for sludge thickening on the 48" basket centrifuge.

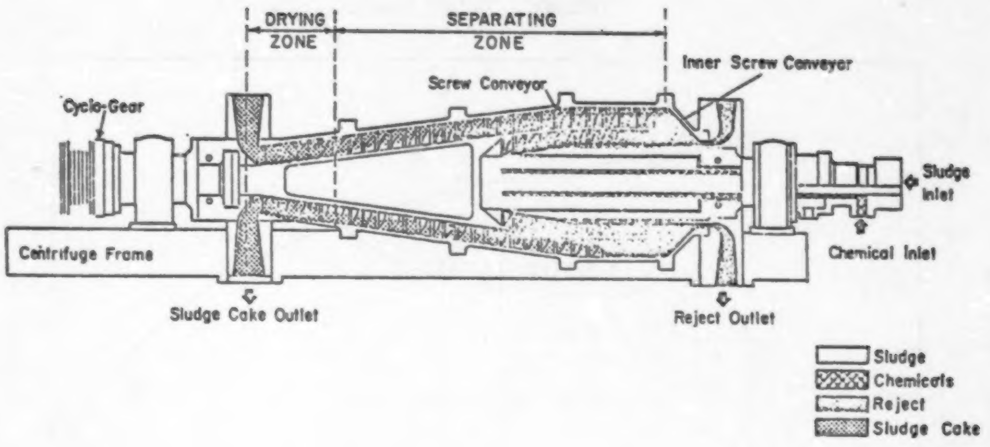
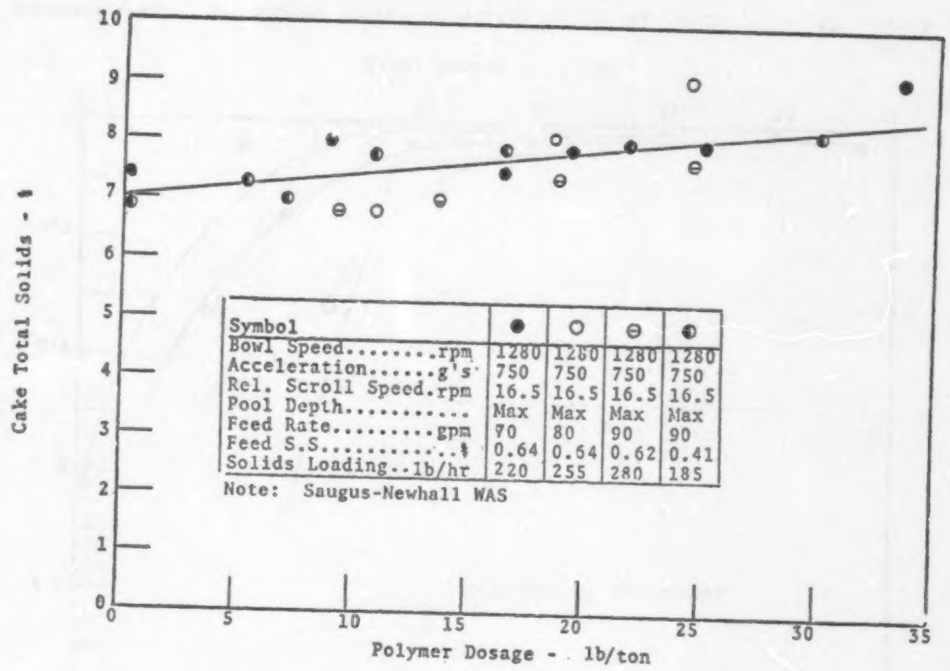

DRYING ZONE
SEPARATING ZONE
Inner Screw Conveyor
Screw Conveyor
Cyclo-Gear
Sludge Inlet
Chemical Inlet
Centrifuge Frame
Sludge Cake Outlet
Reject Outlet
Sludge
Chemicals
Reject
Sludge Cake

Figure 22. Tapered bowl scroll centrifuge



Symbol	●	○	⊖	◐
Bowl Speed........rpm	1280	1280	1280	1280
Acceleration......g's	750	750	750	750
Rel. Scroll Speed.rpm	16.5	16.5	16.5	16.5
Pool Depth...........	Max	Max	Max	Max
Feed Rate.........gpm	70	80	90	90
Feed S.S............%	0.64	0.64	0.62	0.41
Solids Loading..lb/hr	220	255	280	185

Note: Saugus-Newhall WAS

Figure 23. Cake solids vs. polymer dosage for sludge thickening on the 32" x 100" scroll centrifuge.

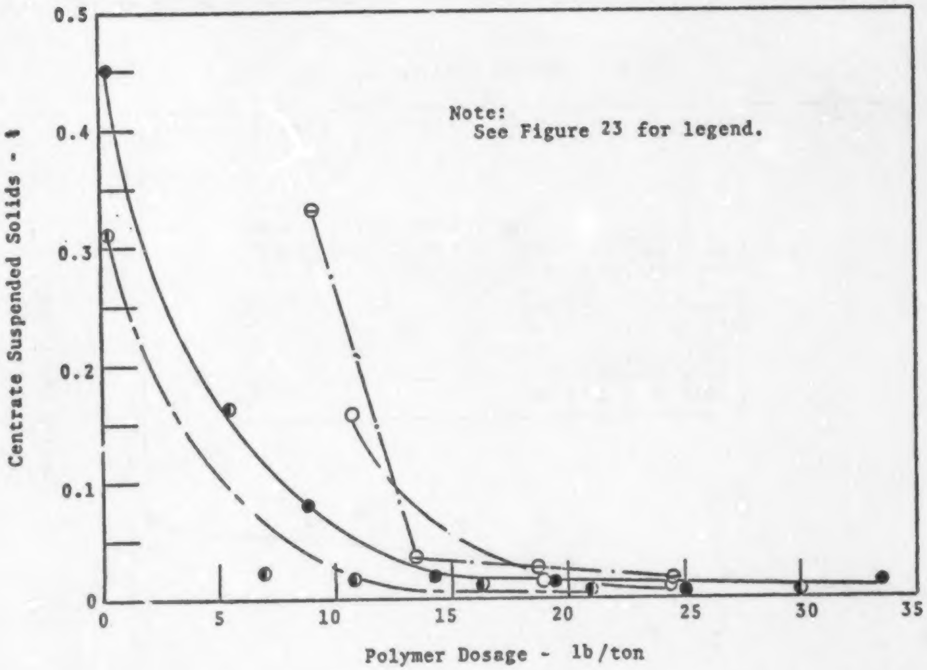


Figure 24. Centrate quality vs. polymer dosage for sludge thickening on the 32" x 100" scroll centrifuge.

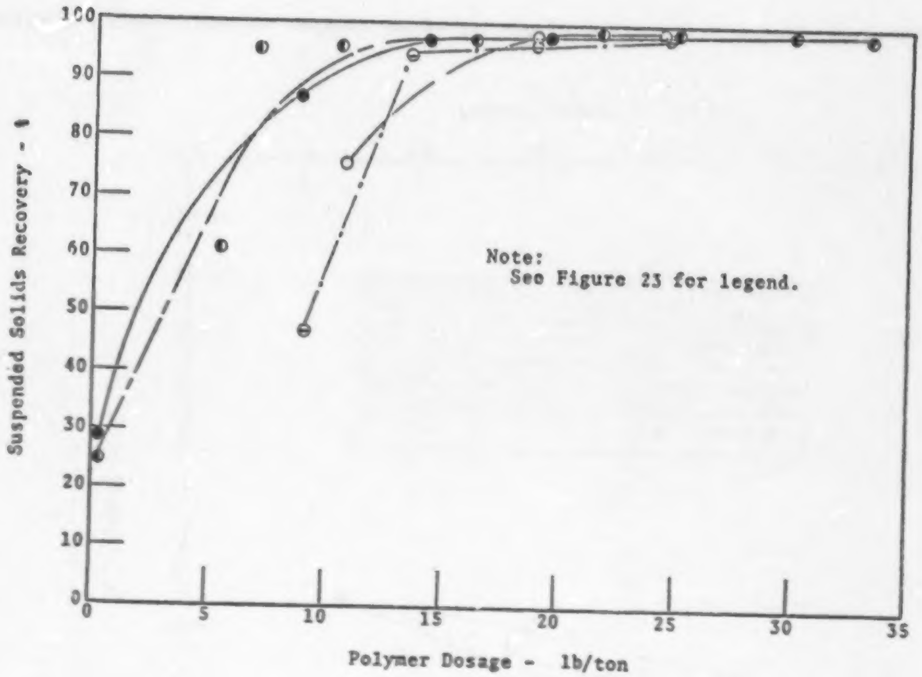
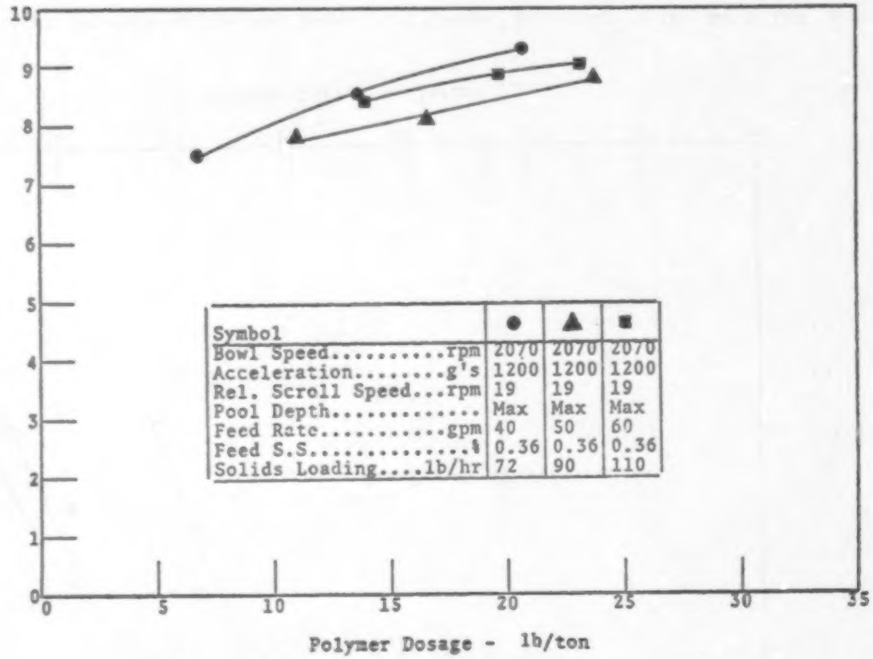


Figure 25. Suspended solids recovery vs. polymer dosage for sludge thickening on the 32" x 100" scroll centrifuge.



Symbol	●	▲	■
Bowl Speed..........rpm	2070	2070	2070
Acceleration.........g's	1200	1200	1200
Rel. Scroll Speed...rpm	19	19	19
Pool Depth.............	Max	Max	Max
Feed Rate...........gpm	40	50	60
Feed S.S...............%	0.36	0.36	0.36
Solids Loading....lb/hr	72	90	110

Figure 26. Cake solids vs. polymer dosage for sludge thickening on the 20" x 62" scroll centrifuge.

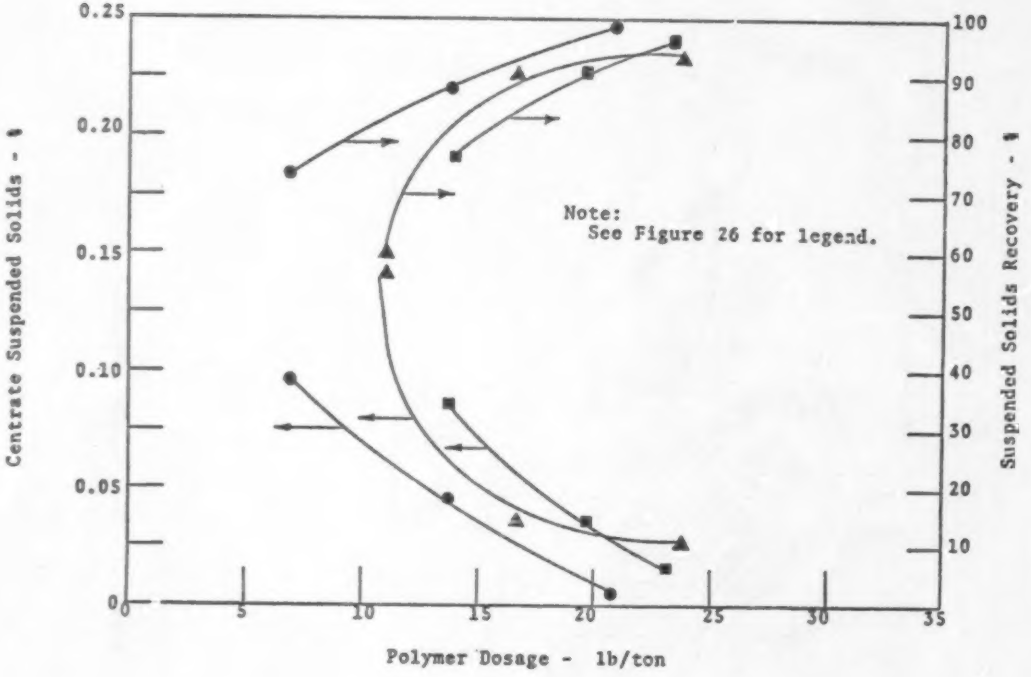


Figure 27. Centrate quality and % removal vs. polymer dosage for sludge thickening on the 20" x 62" scroll centrifuge.

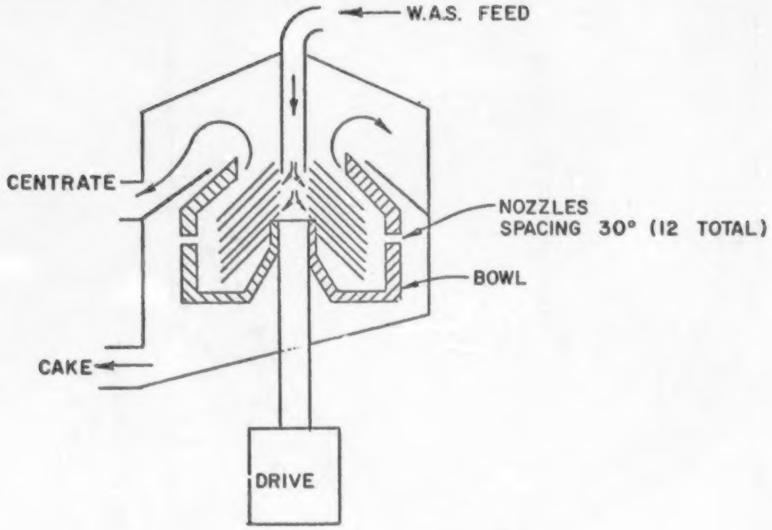
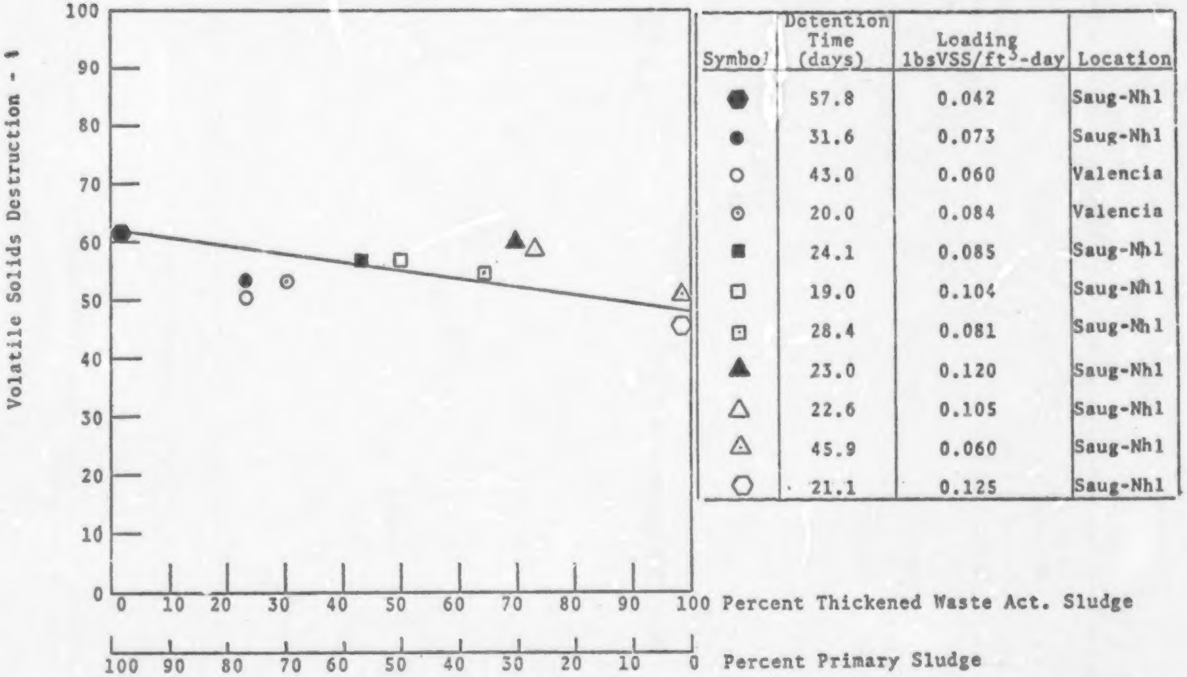

W.A.S. FEED
CENTRATE
NOZZLES
SPACING 30° (12 TOTAL)
BOWL
CAKE
DRIVE

Figure 28. Disc-Nozzle Centrifuge.



Symbol	Detention Time (days)	Loading lbsVSS/ft^3-day	Location
⬢	57.8	0.042	Saug-Nhl
●	31.6	0.073	Saug-Nhl
○	43.0	0.060	Valencia
⊙	20.0	0.084	Valencia
■	24.1	0.085	Saug-Nhl
□	19.0	0.104	Saug-Nhl
⊡	28.4	0.081	Saug-Nhl
▲	23.0	0.120	Saug-Nhl
△	22.6	0.105	Saug-Nhl
◬	45.9	0.060	Saug-Nhl
⬡	21.1	0.125	Saug-Nhl

Figure 29. Anaerobic digestion volatile solids destruction vs. waste activated sludge/primary sludge ratio.

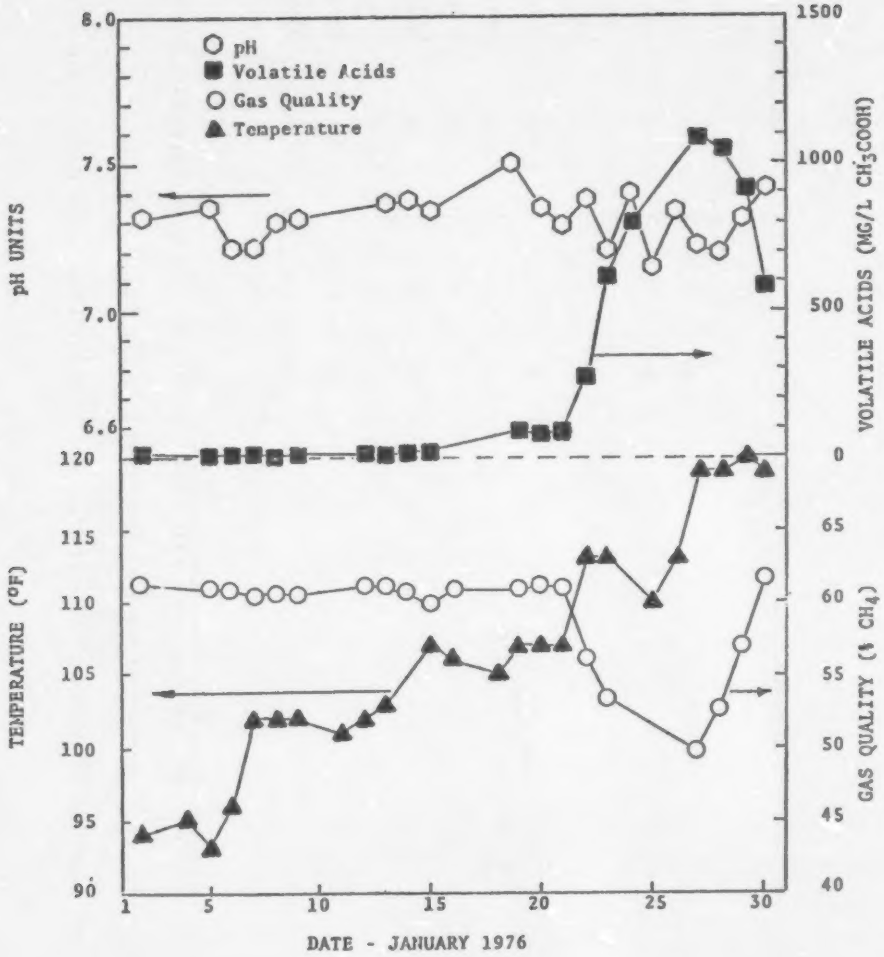


Figure 30. Digester response to change from mesophilic to thermophilic temperature ranges at the JWPCP.

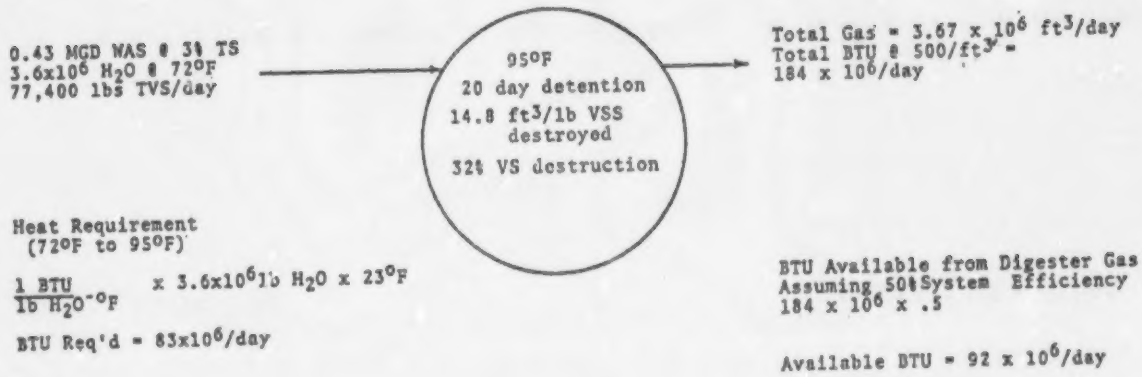
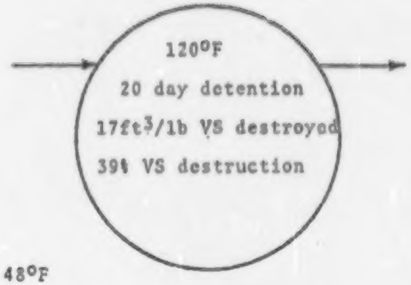

0.43 MGD WAS @ 3% TS
3.6×10^6 H_2O @ 72°F
77,400 lbs TVS/day
95°F
20 day detention
14.8 ft^3/lb VSS destroyed
32% VS destruction
Total Gas = 3.67×10^6 ft^3/day
Total BTU @ 500/ft^3 =
184×10^6/day
Heat Requirement
(72°F to 95°F)
$\frac{1 \text{ BTU}}{\text{lb } H_2O\text{-}^\circ F} \times 3.6 \times 10^6 \text{ lb } H_2O \times 23^\circ F$
BTU Req'd = 83×10^6/day
BTU Available from Digester Gas
Assuming 50% System Efficiency
$184 \times 10^6 \times .5$
Available BTU = 92×10^6/day

Figure 31. Thermal requirements for mesophilic digestion for 100 MGD of secondary treatment.

0.43 MGD WAS @ 3% TS
3.6x10^6 lbs H_2O @ 72°F
77,400 lbs TVS/day


120°F
20 day detention
17ft^3/lb VS destroyed
39% VS destruction

Total Gas = 5.13 x 10^6 ft^3/day
Total BTU @ 500/ft^3 =
256 x 10^6/day

Heat Requirement
(72°F to 120°F)

$\frac{1 \text{ BTU}}{\text{lb } H_2O \text{ °F}}$ x 3.6x10^{6}lb H_2O x 48°F

BTU req'd = 173x10^6/day

BTU Available from Digester Gas
Assuming 50% System Efficiency
256 x 10^6 x .5

Available BTU = 128 x 10^6/day

Figure 32. Thermal requirements for thermophilic digestion of the WAS from 100 MGD of secondary treatment.

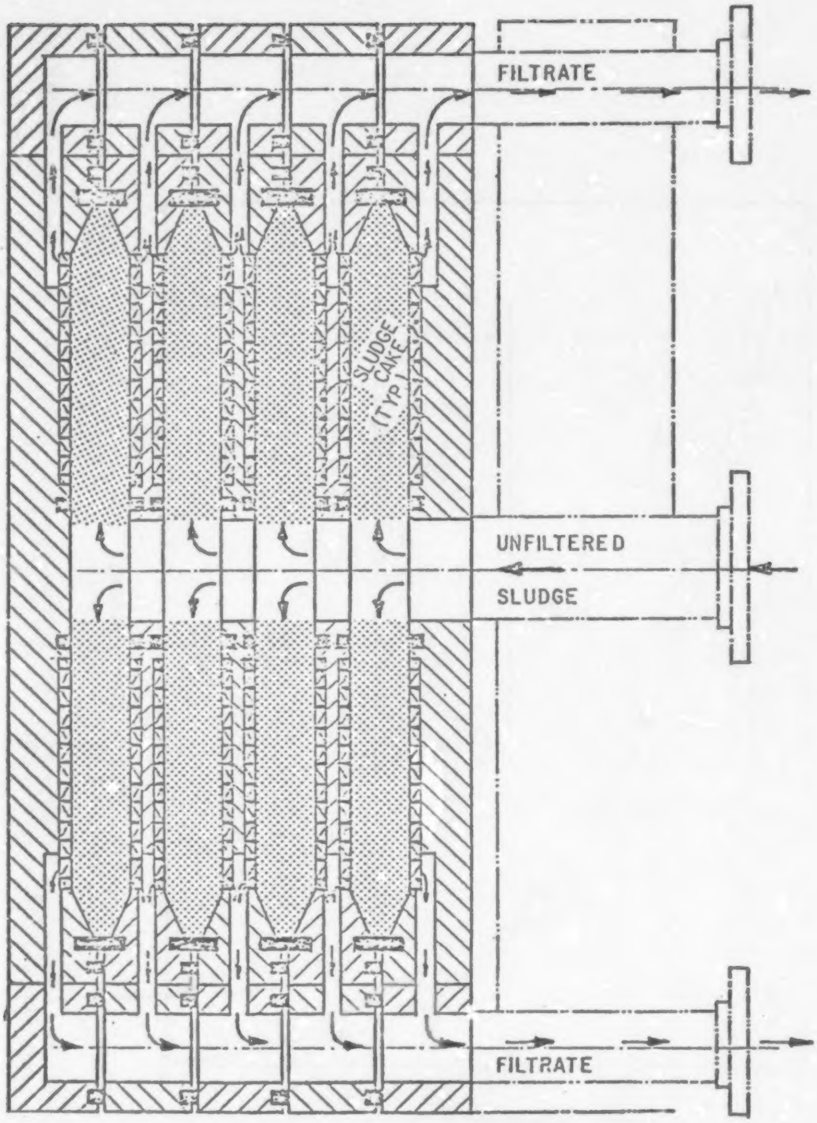
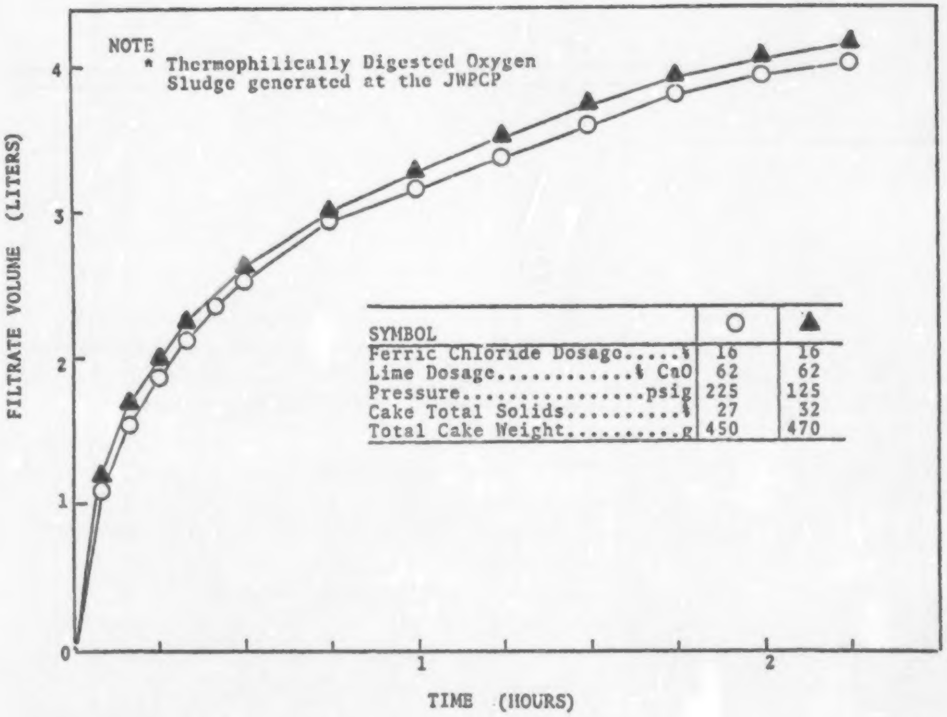

FILTRATE
SLUDGE CAKE (TYP)
UNFILTERED
SLUDGE
FILTRATE

Figure 33. Filter press.



SYMBOL	○	▲
Ferric Chloride Dosage.....%	16	16
Lime Dosage............% CaO	62	62
Pressure................psig	225	125
Cake Total Solids..........%	27	32
Total Cake Weight..........g	450	470

Figure 34. Effects of operating pressure on filter performance on filtrate volume vs. time relationship for dewatering digested oxygen sludge* on the 0.33 ft^2 pressure filter.

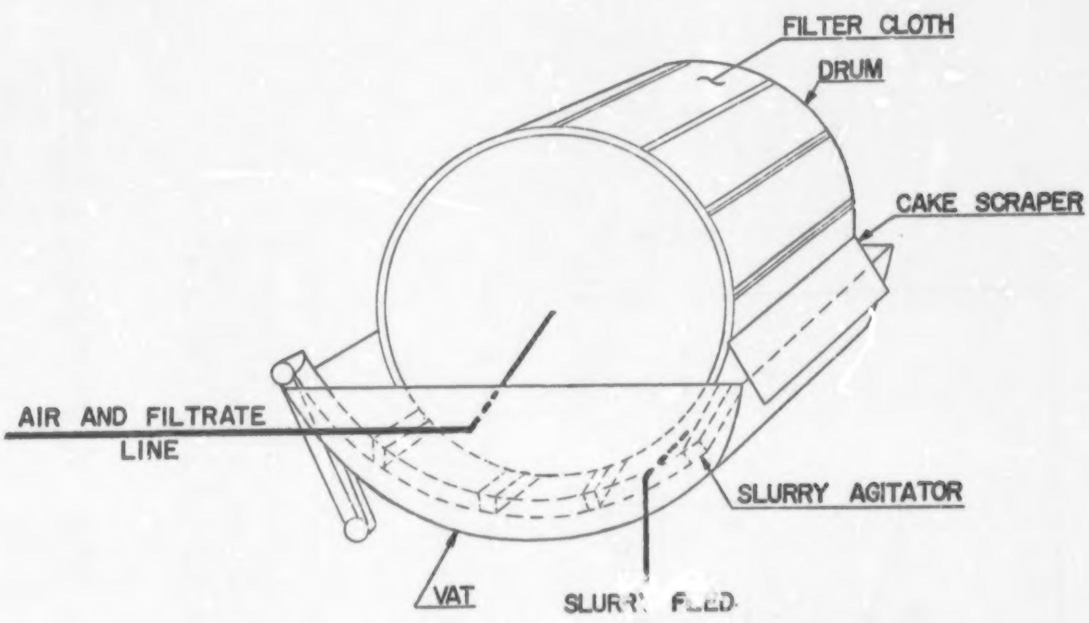
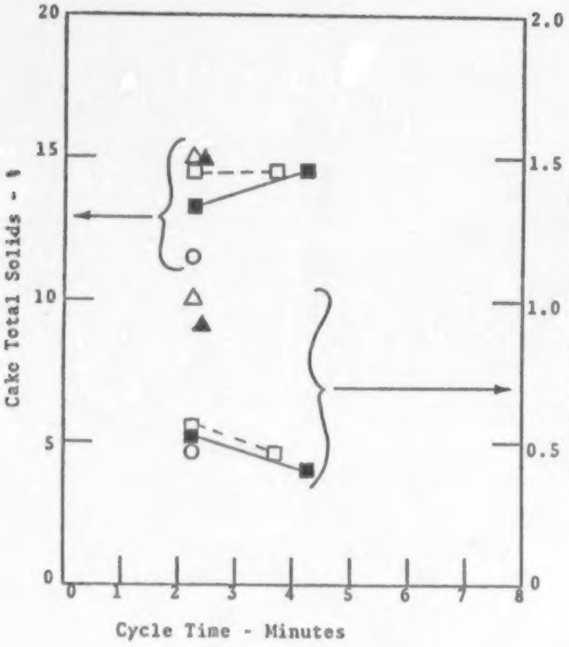

FILTER CLOTH
DRUM
CAKE SCRAPER
AIR AND FILTRATE LINE
SLURRY AGITATOR
VAT
SLURRY FEED

Figure 35. Rotary Drum Vacuum Filter.



Symbol					
Cloth Type.............	Nylon	Nylon	Nylon	Nylon	Nylon
Vacuum............in.Hg	22	22	22	22	22
Feed S. S.............%	1.02	1.02	1.02	1.02	1.02
Chemical Type..........	None	$FeCl_3$	$FeCl_3$	$FeCl_3$	$FeCl_3$
Chemical Dosage..lb/ton	----	75	125	200	300

Figure 36. Cake solids and filter yield vs. cycle time for dewatering aerobically digested waste activated sludge on the 3' x 1' rotary drum vacuum filter.

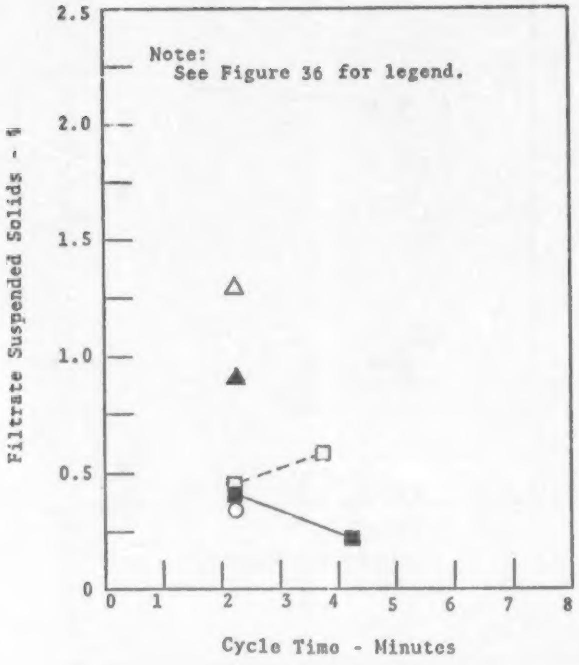
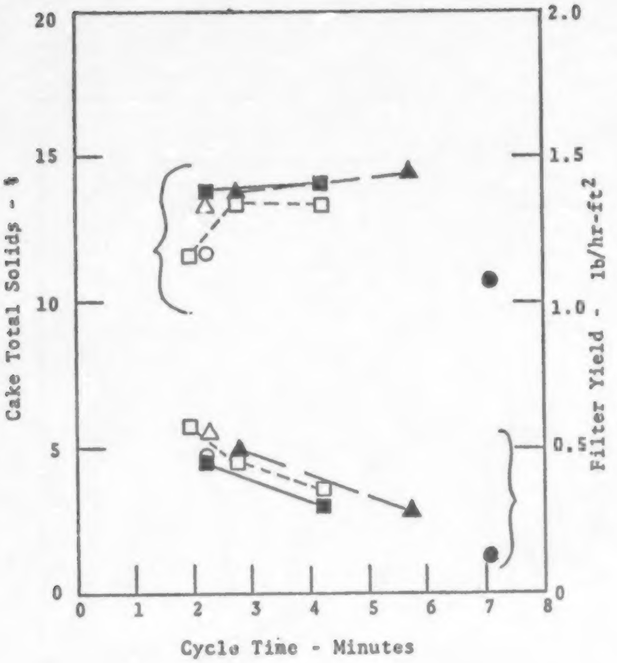


Figure 37. Filtrate quality vs. cycle time for dewatering aerobically digested waste activated sludge on the 3' x 1' rotary drum vacuum filter.



Symbol						
Cloth Type............	Nylon	Nylon	Nylon	Nylon	Nylon	Nylon
Vacuum...........in.Hg	22	22	21.5	21.5	22	22
Feed S. S.............	1.02	1.02	1.02	1.02	1.02	1.02
Chemical Type.........	None	None	Lime	Lime	Lime	Lime
Chemical Dosage.lb/ton	0	0	100	200	300	600

Figure 38. Cake solids and yield vs. cycle time for dewatering aerobically digested waste activated sludge on the 3' x 1' rotary drum vacuum filter.

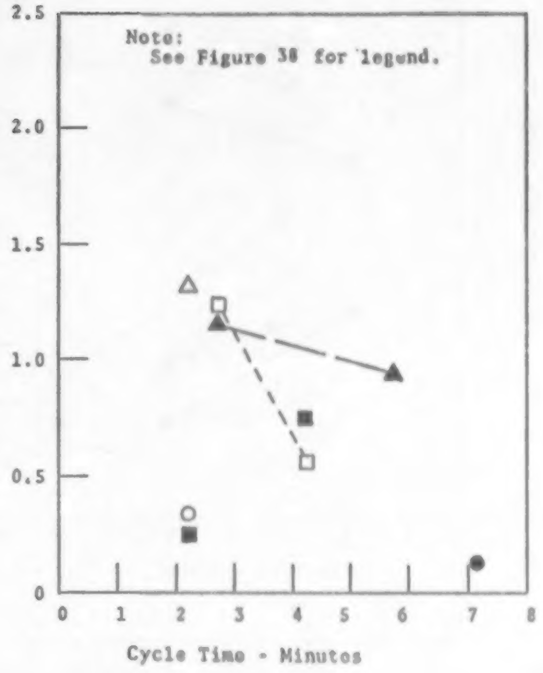
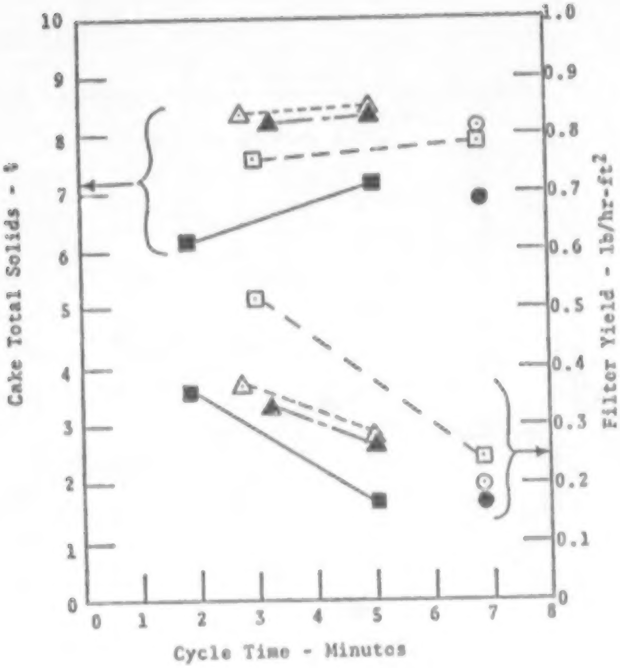


Figure 39. Filtrate quality vs. cycle time for dewatering aerobically digested waste activated sludge on the 3' x 1' rotary drum vacuum filter.



Symbol						
Cloth Type.............	Nylon	Nylon	Nylon	Nylon	Nylon	Nylon
Vacuum............in.Hg	22	22	21	14	22	22
Feed S. S..............%	1.97	1.97	1.97	1.97	1.97	1.97
Chemical...............	None	Polymer	$FeCl_3$	$FeCl_3$	Lime	Lime
Chemical Dosage.lb/ton	----	5	50	125	100	200

Note 1. 43% Waste Activated Sludge-57% Primary Sludge

Figure 40. Cake solids and yield vs. cycle time for dewatering digested blend[1] on the 3' x 1' rotary drum vacuum filter.

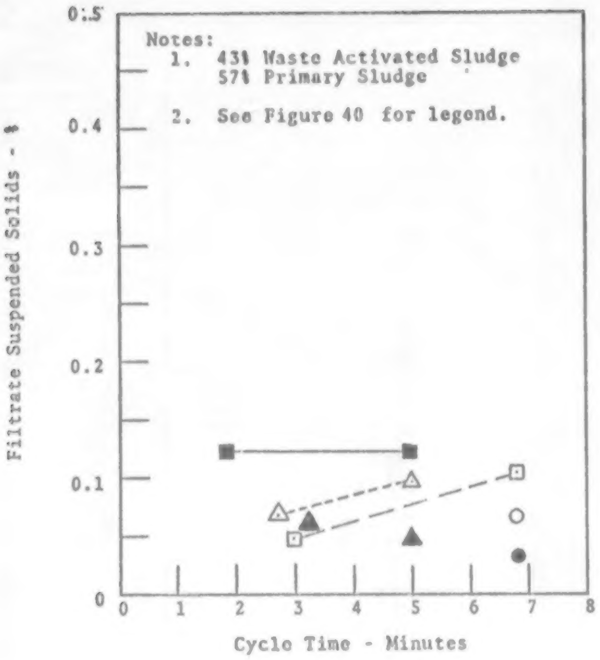


Figure 41. Filtrate quality vs. cycle time for dewatering digested blend[1] on the 3' x 1' rotary drum vacuum filter.

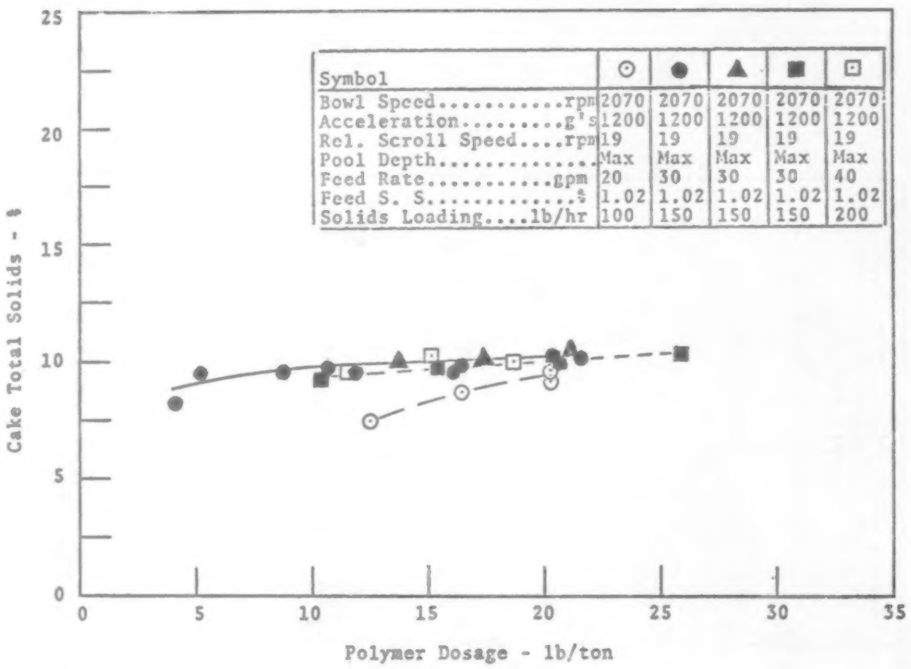


Figure 42. Cake solids vs. polymer dosage for dewatering aerobically digested waste activated sludge on the 20" x 62" scroll centrifuge.

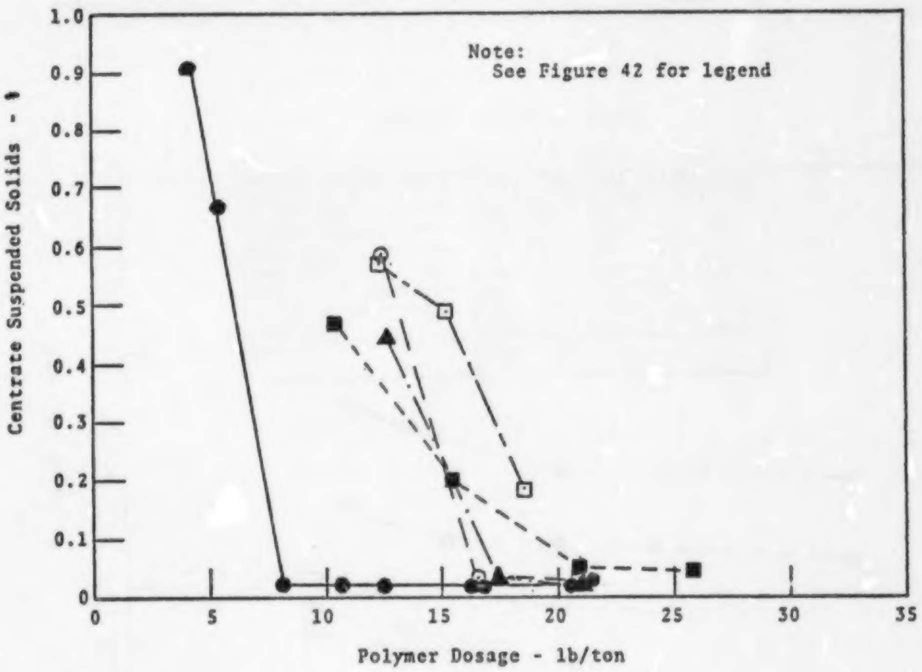


Figure 43. Centrate quality vs. polymer dosage for dewatering aerobically digested waste activated sludge on the 20" x 62" scroll centrifuge.

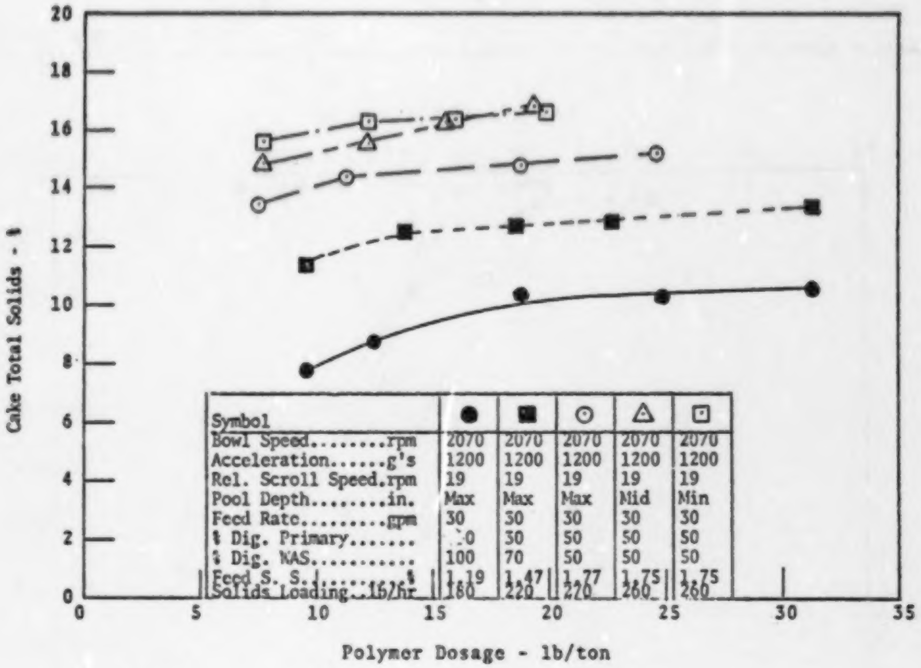
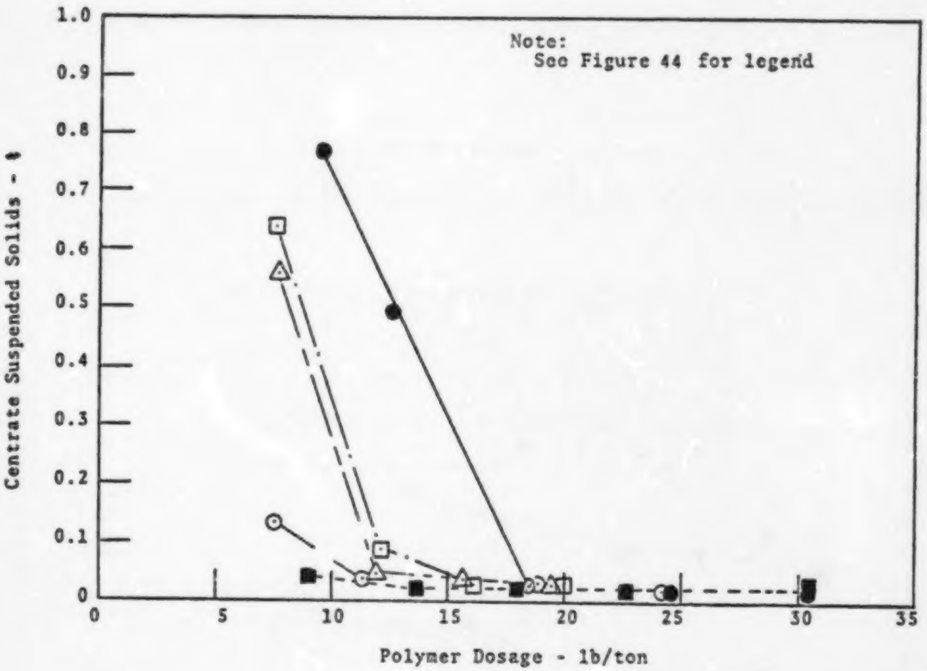
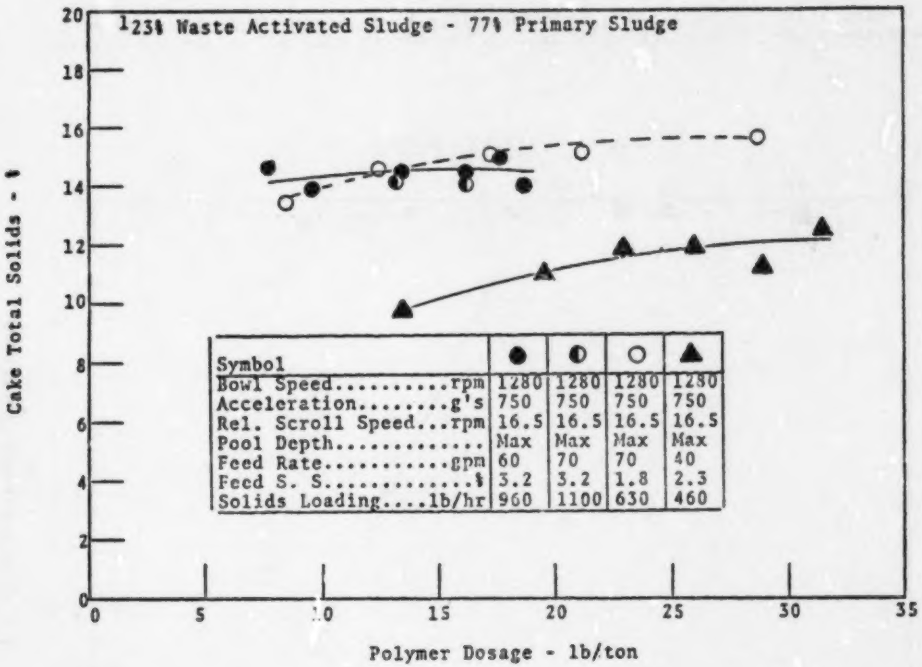


Figure 44. Cake solids vs. polymer dosage for dewatering aerobically digested waste activated sludge and anaerobically digested primary sludge on the 20" x 62" scroll centrifuge.



Note:
See Figure 44 for legend

Figure 45. Centrate quality vs. polymer dosage for dewatering aerobically digested waste activated sludge and anaerobically digested primary sludge on the 20" x 62" scroll centrifuge.



Symbol	●	◐	○	▲
Bowl Speed..........rpm	1280	1280	1280	1280
Acceleration........g's	750	750	750	750
Rel. Scroll Speed...rpm	16.5	16.5	16.5	16.5
Pool Depth............	Max	Max	Max	Max
Feed Rate...........gpm	60	70	70	40
Feed S. S.............%	3.2	3.2	1.8	2.3
Solids Loading....lb/hr	960	1100	630	460

Figure 46. Cake solids vs. polymer dosage for digested blend[1] dewatering on the 32" x 100" scroll centrifuge

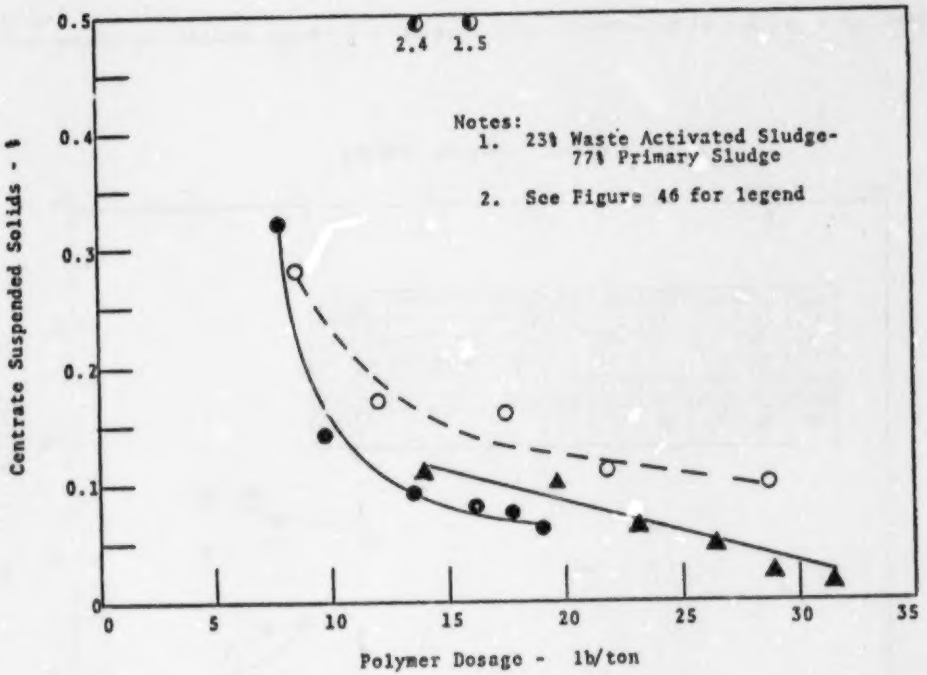
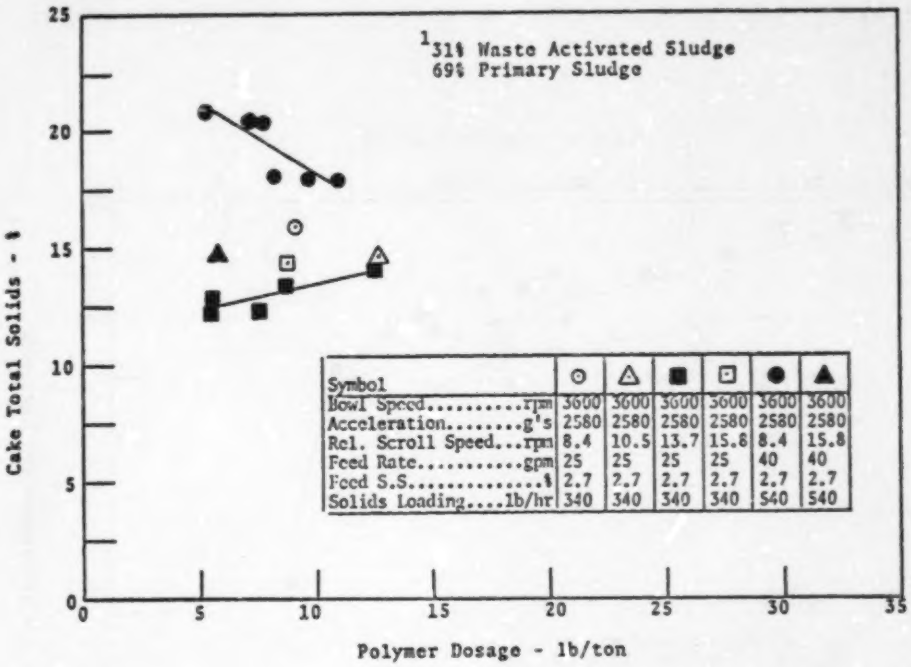


Figure 47. Centrate quality vs. polymer dosage for digested blend[1] dewatering on the 32" x 100" scroll centrifuge.



Symbol	⊙	△	■	⊡	●	▲
Bowl Speed..........rpm	3600	3600	3600	3600	3600	3600
Acceleration........g's	2580	2580	2580	2580	2580	2580
Rel. Scroll Speed...rpm	8.4	10.5	13.7	15.8	8.4	15.8
Feed Rate...........gpm	25	25	25	25	40	40
Feed S.S..............%	2.7	2.7	2.7	2.7	2.7	2.7
Solids Loading....lb/hr	340	340	340	340	540	540

Figure 48. Cake solids vs. polymer dosage for digested blend¹ dewatering on the 14" x 48" scroll centrifuge.

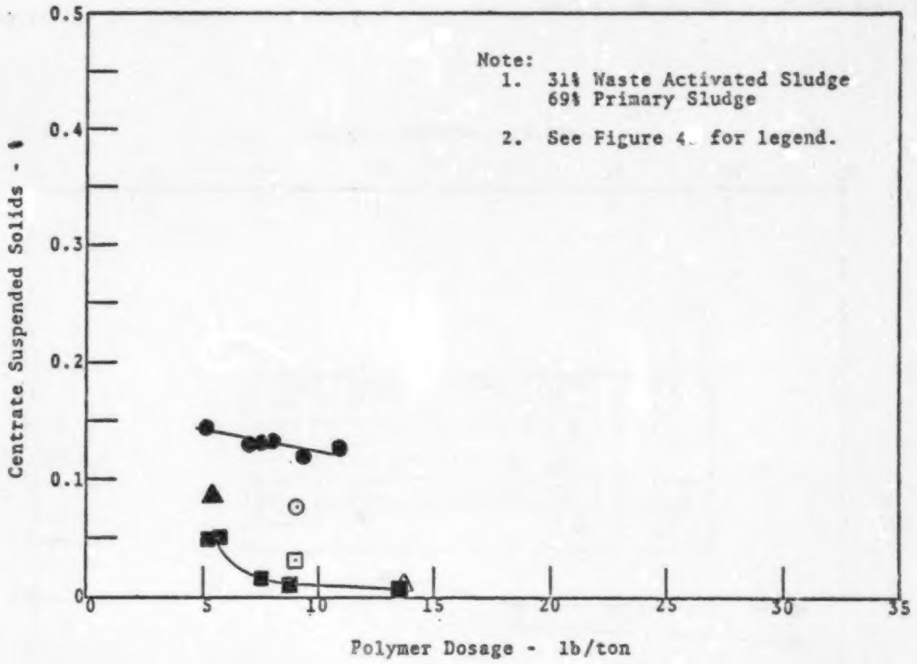


Figure 49. Centrate quality vs. polymer dosage for digested blend^1 dewatering on the 14" x 48" scroll centrifuge.

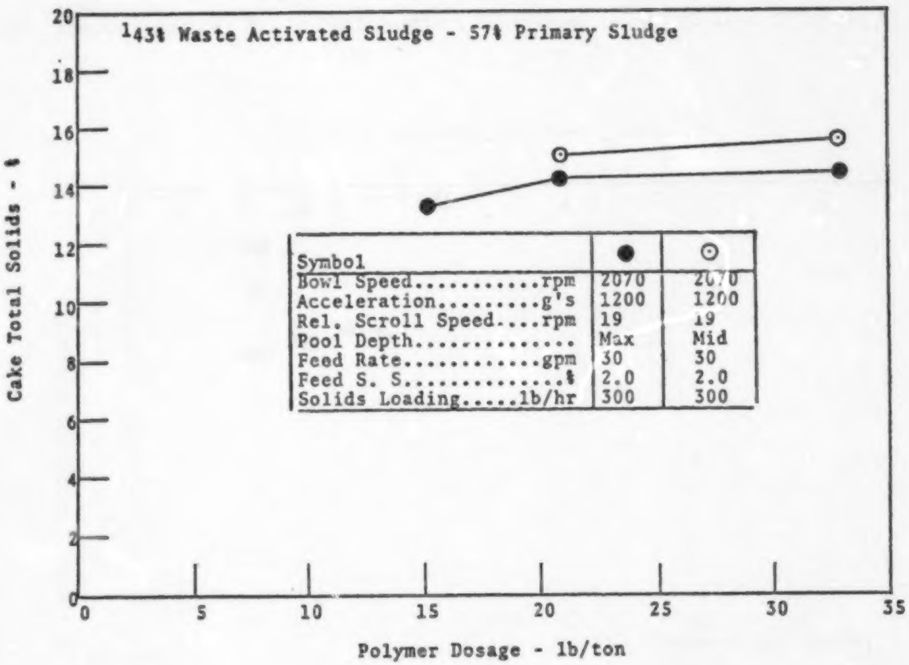


Figure 50. Cake solids vs. polymer dosage for digested blend[1] dewatering on the 20" x 62" scroll centrifuge.

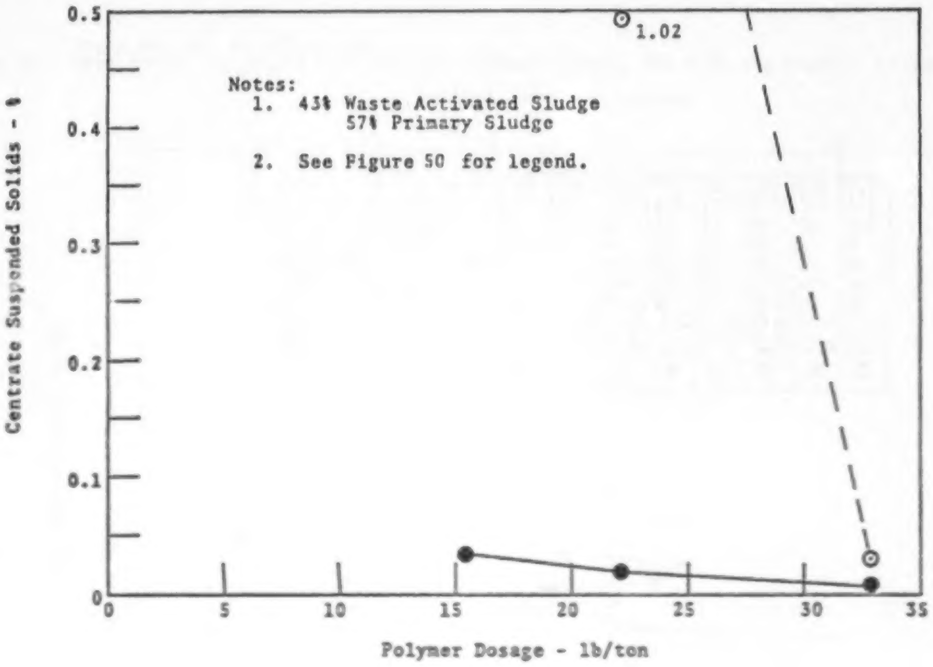
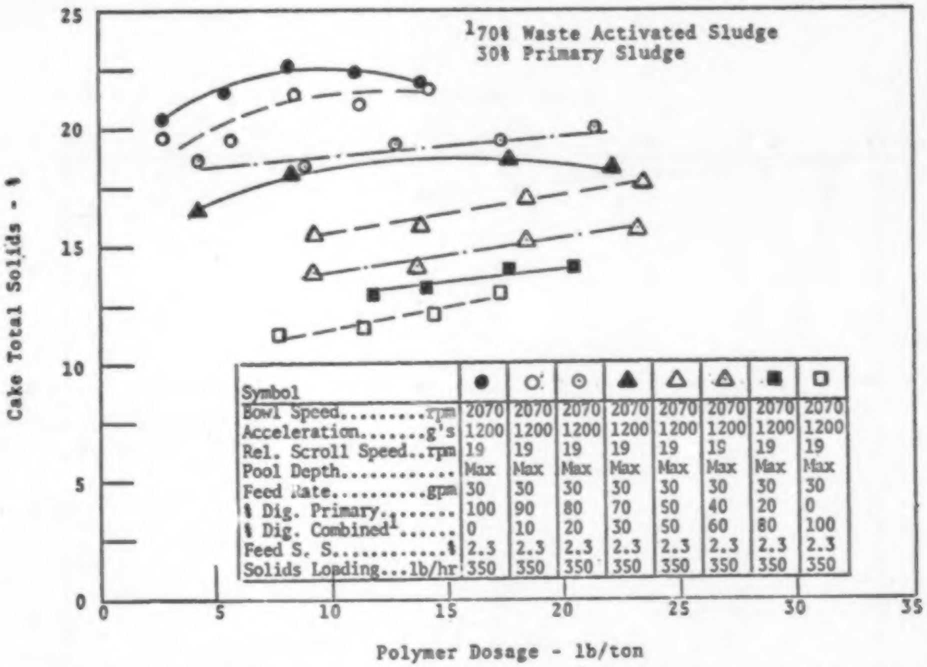


Figure 51. Centrate quality vs. polymer dosage for digested blend[1] dewatering on the 20" x 62" scroll centrifuge.



Symbol	●	○	⊙	▲	△	◬	■	□
Bowl Speed..........rpm	2070	2070	2070	2070	2070	2070	2070	2070
Acceleration........g's	1200	1200	1200	1200	1200	1200	1200	1200
Rel. Scroll Speed..rpm	19	19	19	19	19	19	19	19
Pool Depth............	Max	Max	Max	Max	Max	Max	Max	Max
Feed Rate..........gpm	30	30	30	30	30	30	30	30
% Dig. Primary.........	100	90	80	70	50	40	20	0
% Dig. Combined¹......	0	10	20	30	50	60	80	100
Feed S. S.............%	2.3	2.3	2.3	2.3	2.3	2.3	2.3	2.3
Solids Loading...lb/hr	350	350	350	350	350	350	350	350

Figure 52. Cake solids vs. polymer dosage for digested blend¹ and digested primary sludge dewatering on the 20" x 62" scroll centrifuge.

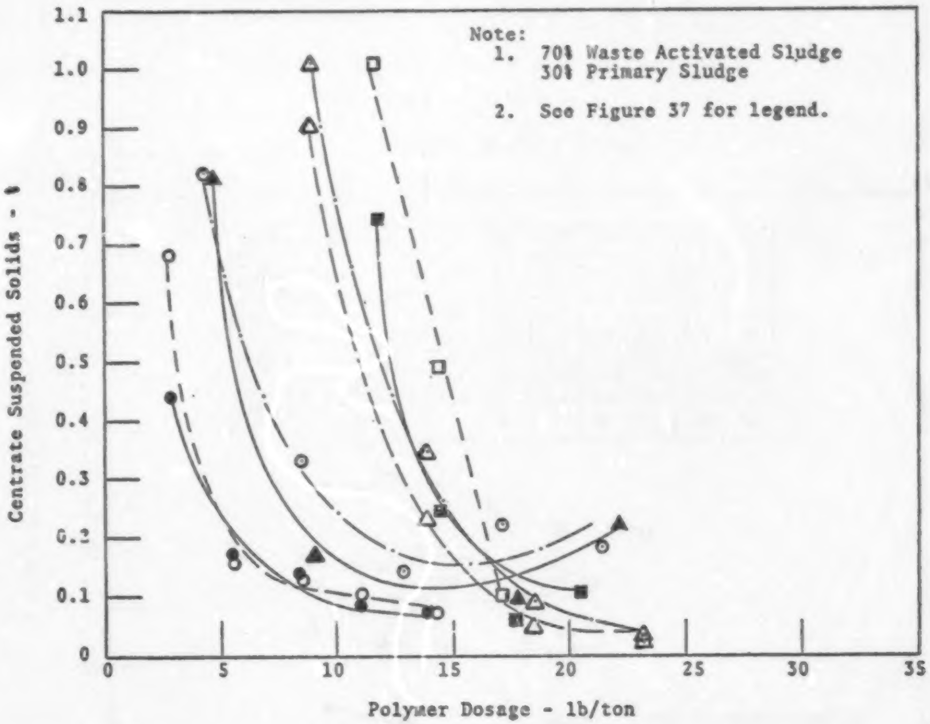
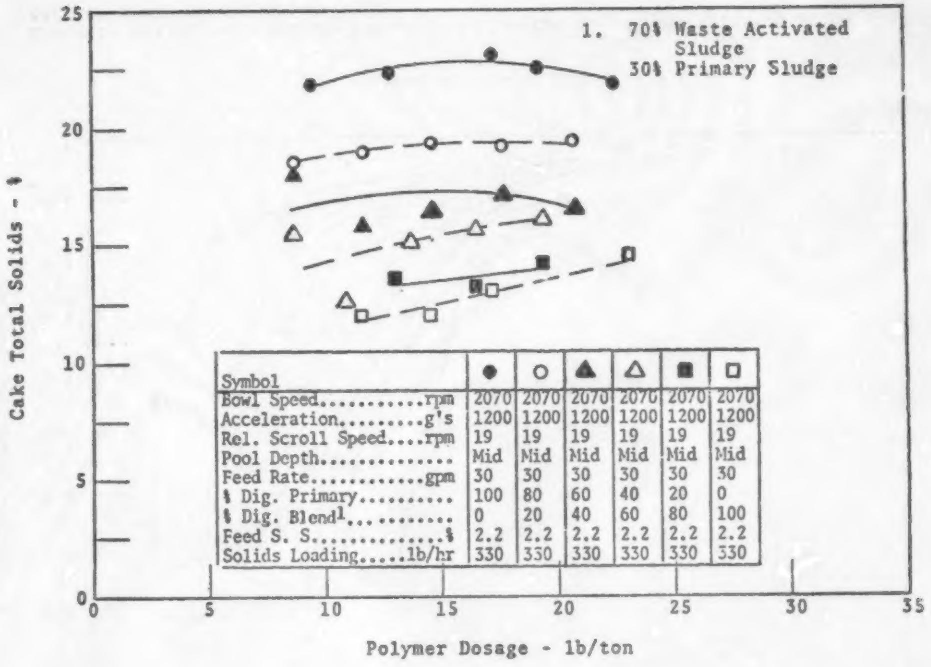


Figure 53. Centrate quality vs. polymer dosage for digested blend[1] and digested primary sludge dewatering on the 20" x 62" scroll centrifuge.



Symbol	●	○	▲	△	■	□
Bowl Speed...........rpm	2070	2070	2070	2070	2070	2070
Acceleration.........g's	1200	1200	1200	1200	1200	1200
Rel. Scroll Speed....rpm	19	19	19	19	19	19
Pool Depth..............	Mid	Mid	Mid	Mid	Mid	Mid
Feed Rate............gpm	30	30	30	30	30	30
% Dig. Primary..........	100	80	60	40	20	0
% Dig. Blend[1]..........	0	20	40	60	80	100
Feed S. S..............%	2.2	2.2	2.2	2.2	2.2	2.2
Solids Loading.....lb/hr	330	330	330	330	330	330

Figure 54. Cake solids vs. polymer dosage for digested blend[1] and digested primary sludge dewatering on the 20" x 62" scroll centrifuge.

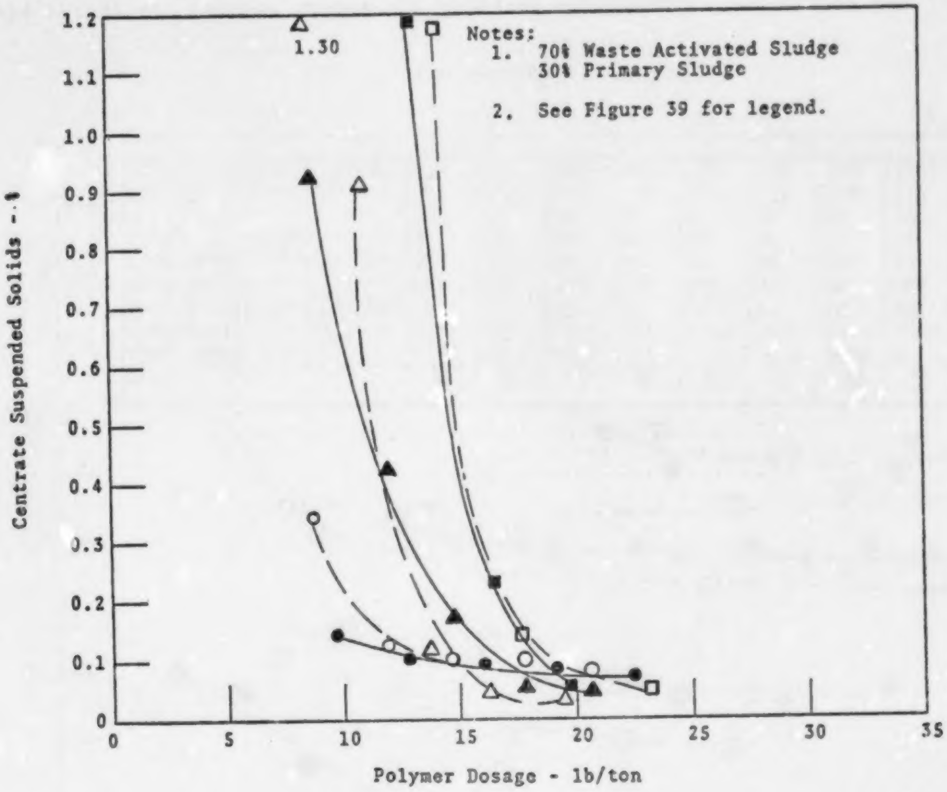
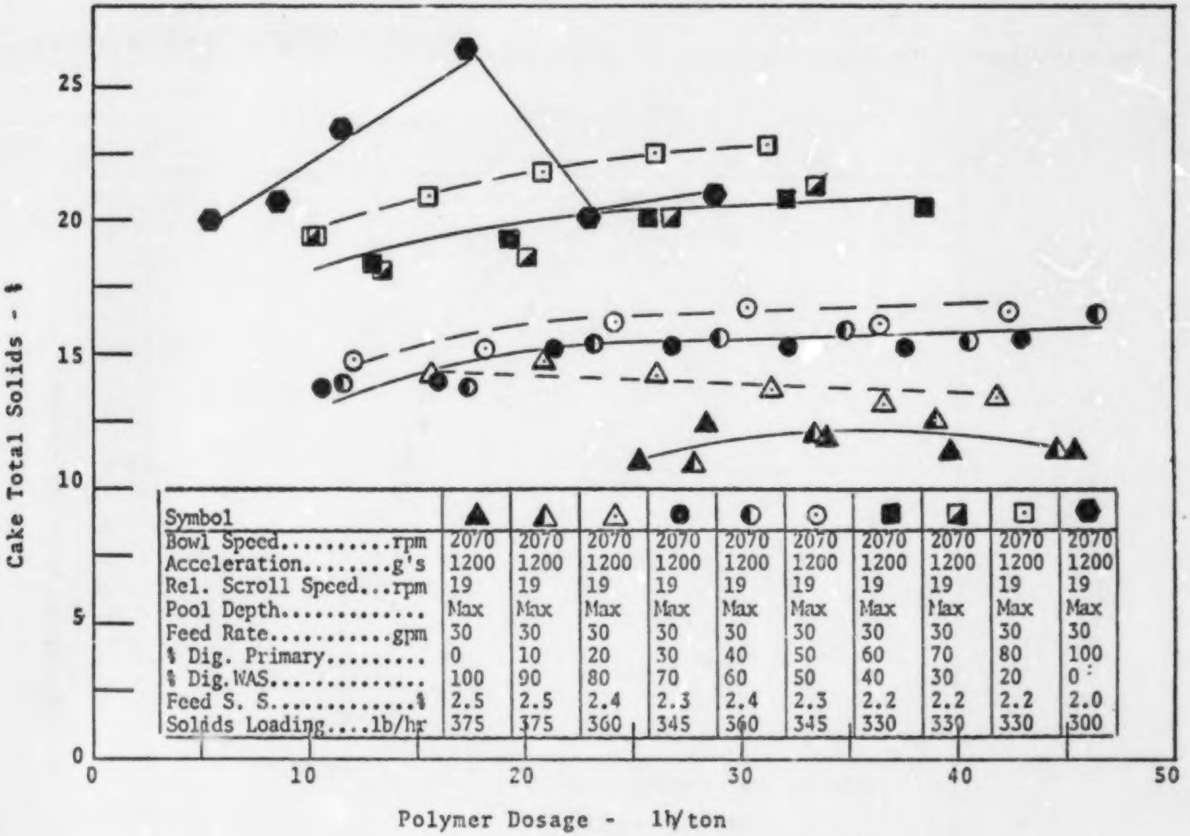


Figure 55. Centrate quality vs. polymer dosage for digested blend[1] and digested primary sludge dewatering on the 20" x 62" scroll centrifuge.



Symbol	▲	◭	△	●	◐	⊙	■	◪	⊡	⬢
Bowl Speed..........rpm	2070	2070	2070	2070	2070	2070	2070	2070	2070	2070
Acceleration........g's	1200	1200	1200	1200	1200	1200	1200	1200	1200	1200
Rel. Scroll Speed...rpm	19	19	19	19	19	19	19	19	19	19
Pool Depth............	Max	Max	Max	Max	Max	Max	Max	Max	Max	Max
Feed Rate...........gpm	30	30	30	30	30	30	30	30	30	30
% Dig. Primary.........	0	10	20	30	40	50	60	70	80	100
% Dig. WAS.............	100	90	80	70	60	50	40	30	20	0
Feed S. S.............%	2.5	2.5	2.4	2.3	2.4	2.3	2.2	2.2	2.2	2.0
Solids Loading....lb/hr	375	375	360	345	360	345	330	330	330	300

Figure 56. Cake solids vs. polymer dosage for digested wasted activated sludge and digested primary sludge dewatering on the 20" x 62" scroll centrifuge.

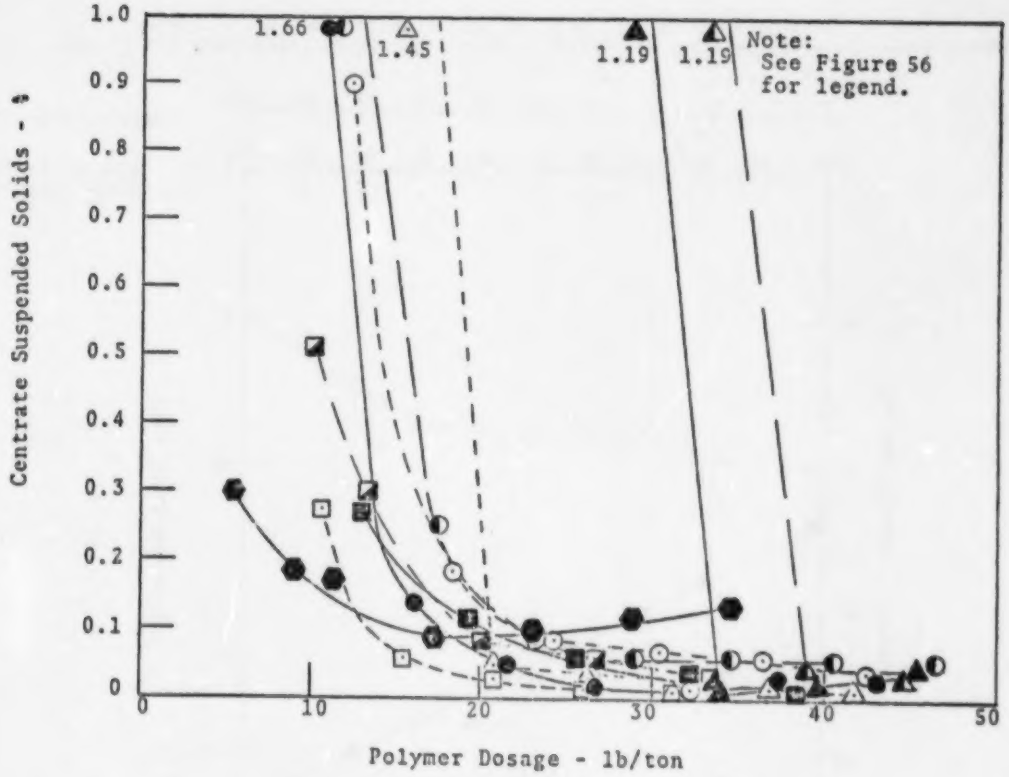


Figure 57. Centrate quality vs. polymer dosage for digested waste activated sludge and digested primary sludge dewatering on the 20" x 62" scroll centrifuge.

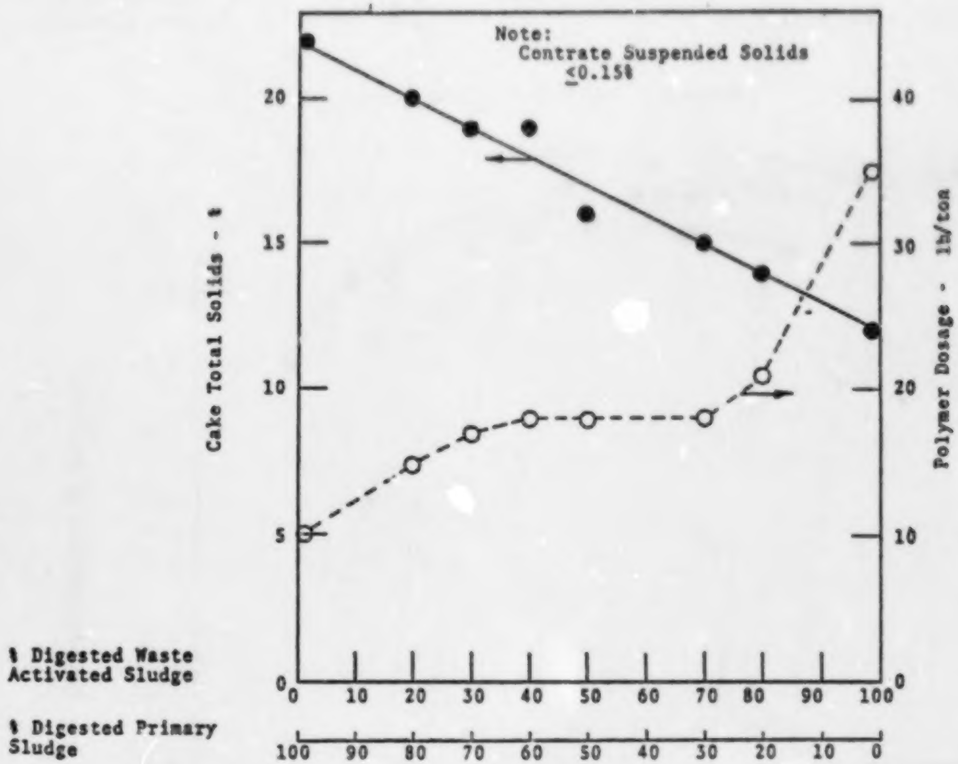
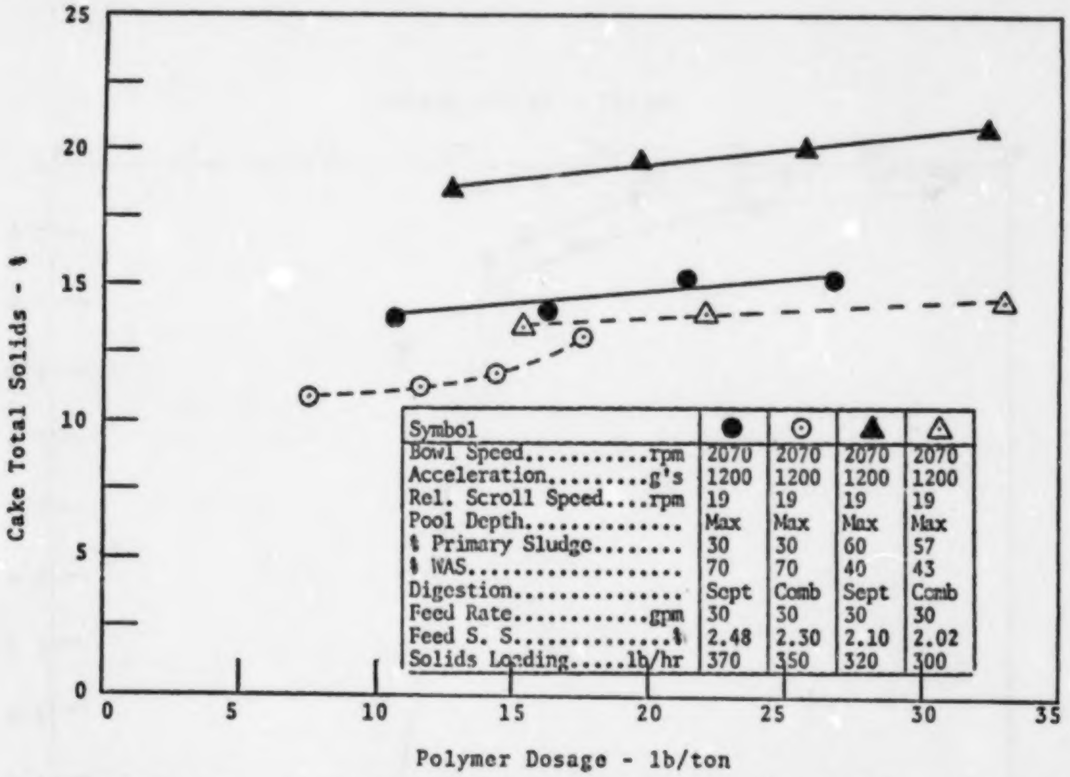


Figure 58. Cake solids and polymer dosage vs. sludge fraction for digested waste activated sludge and digested primary sludge dewatering on the 20" x 62" scroll centrifuge.



Symbol	●	⊙	▲	△
Bowl Speed..........rpm	2070	2070	2070	2070
Acceleration.........g's	1200	1200	1200	1200
Rel. Scroll Speed....rpm	19	19	19	19
Pool Depth..............	Max	Max	Max	Max
% Primary Sludge........	30	30	60	57
% WAS...................	70	70	40	43
Digestion...............	Sept	Comb	Sept	Comb
Feed Rate............gpm	30	30	30	30
Feed S. S...............%	2.48	2.30	2.10	2.02
Solids Loading.....lb/hr	370	350	320	300

Figure 59. Comparison of separate and combined digestion; cake solids vs. polymer dosage for digested sludge dewatering on the 20" x 62" scroll centrifuge.

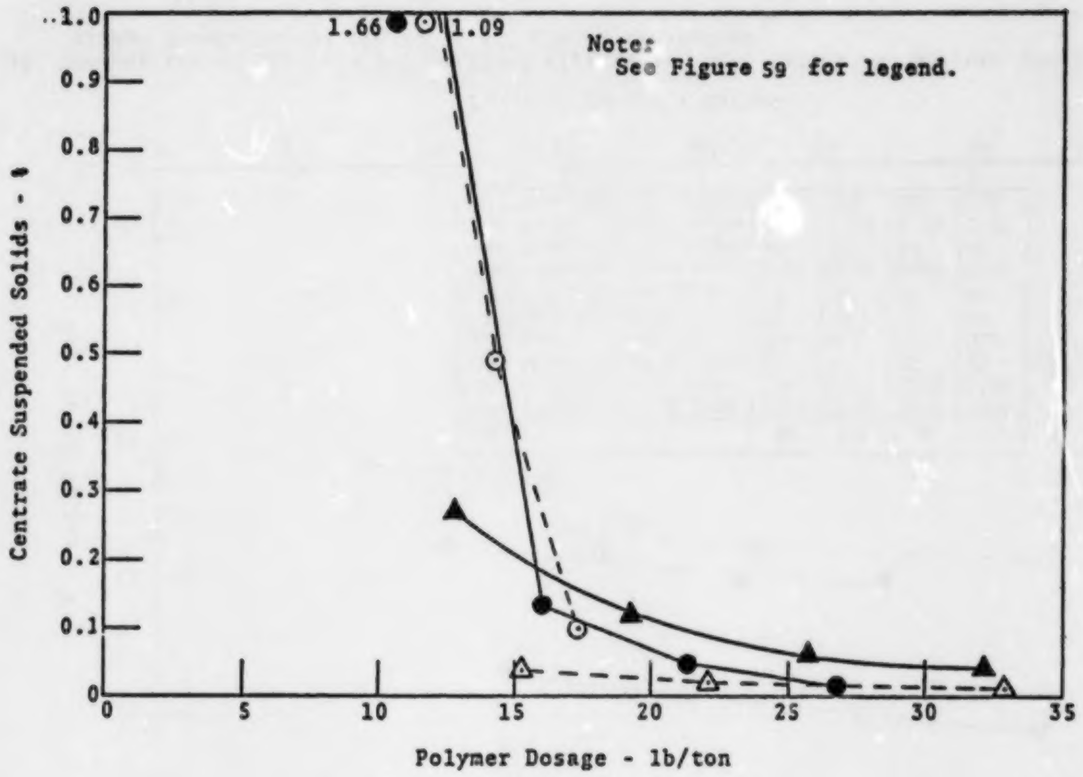
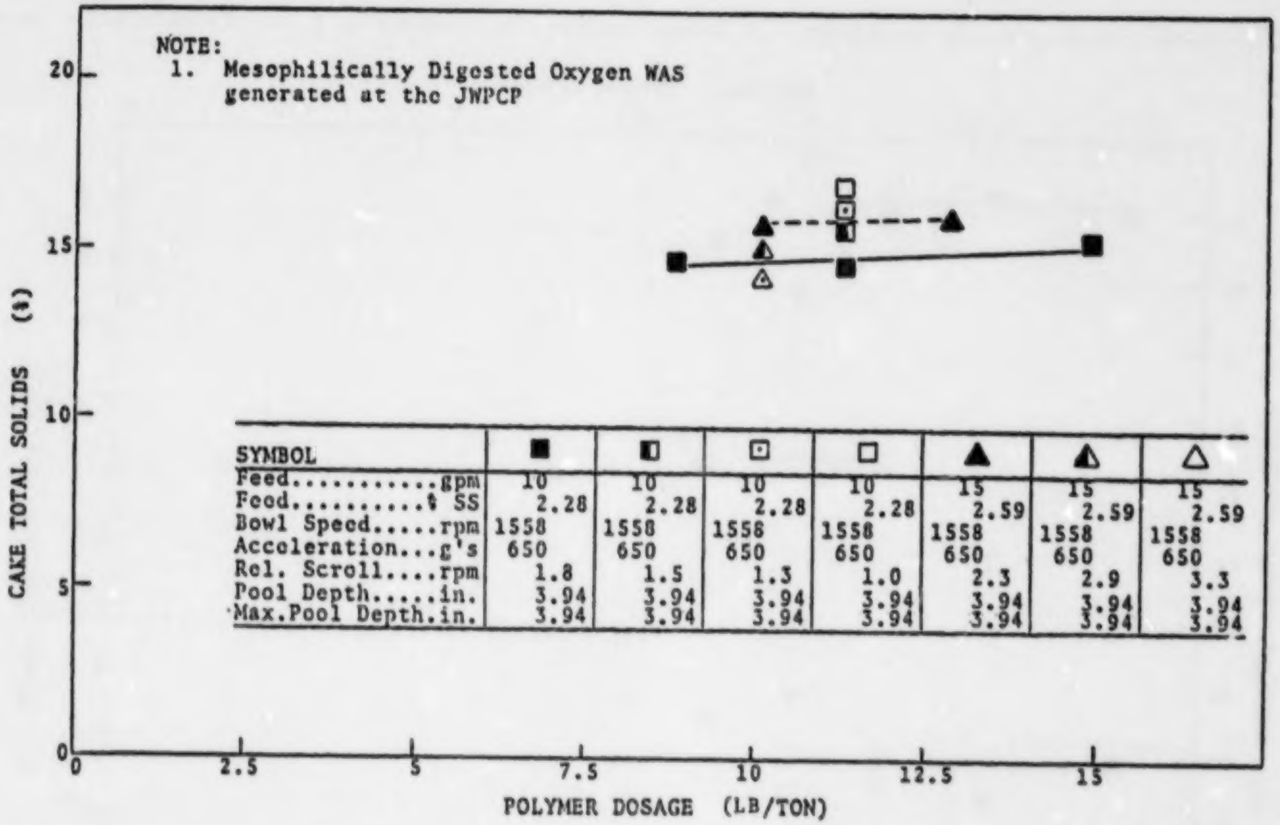


Figure 60. Comparison of separate and combined digestion; centrate quality vs. polymer dosage for digested sludge dewatering on the 20" x 62" scroll centrifuge.


NOTE:
1. Mesophilically Digested Oxygen WAS generated at the JWPCP

SYMBOL	■	◧	⊡	□	▲	◭	△
Feed............gpm	10	10	10	10	15	15	15
Feed..........% SS	2.28	2.28	2.28	2.28	2.59	2.59	2.59
Bowl Speed.....rpm	1558	1558	1558	1558	1558	1558	1558
Acceleration...g's	650	650	650	650	650	650	650
Rel. Scroll....rpm	1.8	1.5	1.3	1.0	2.3	2.9	3.3
Pool Depth.....in.	3.94	3.94	3.94	3.94	3.94	3.94	3.94
Max.Pool Depth.in.	3.94	3.94	3.94	3.94	3.94	3.94	3.94

Figure 61. Cake solids vs. polymer dosage for dewatering digested oxygen sludge[1] on the 18" x 54" scroll centrifuge.

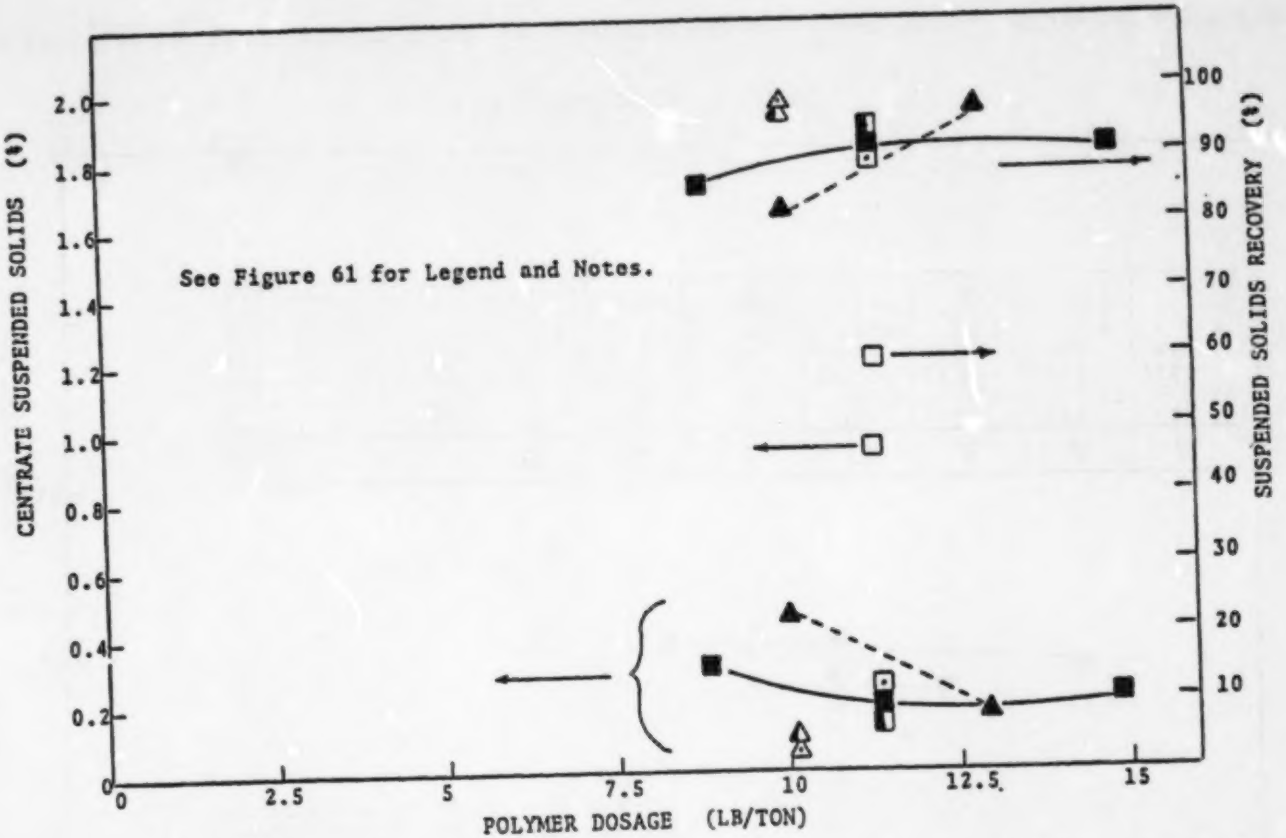
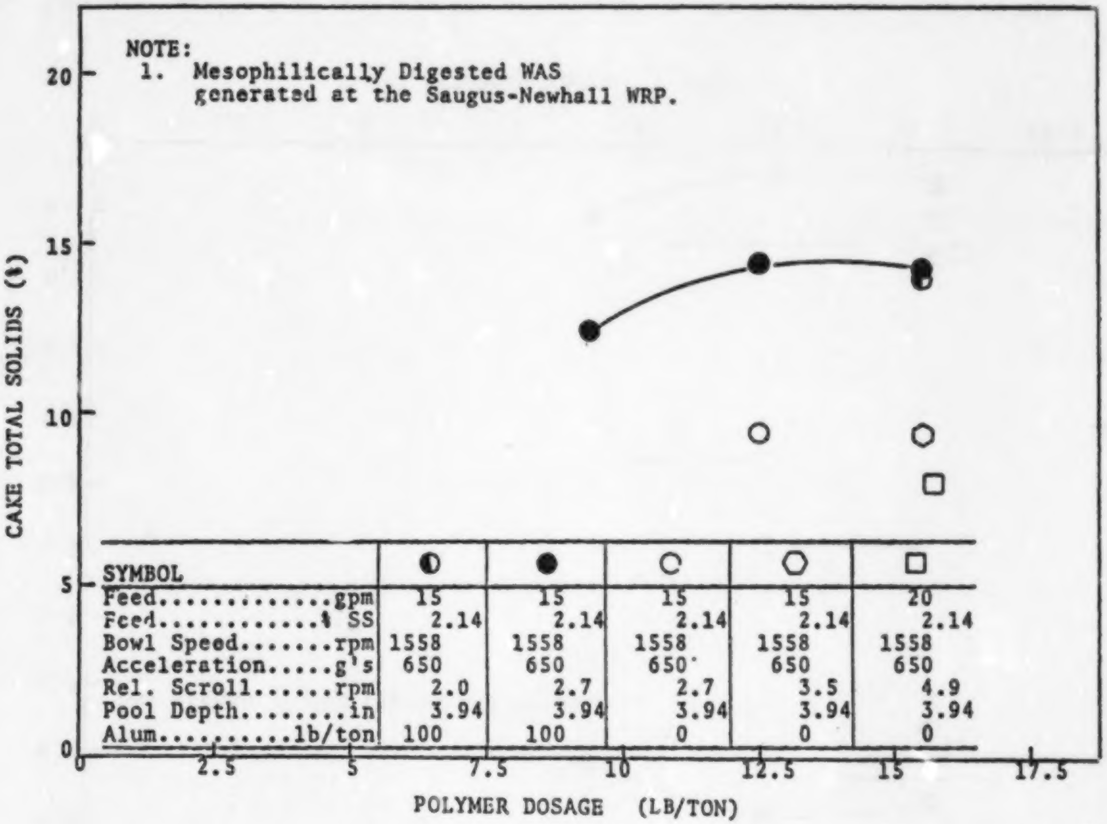


Figure 62. Centrate quality & SS recovery vs. polymer dosage for dewatering digested oxygen waste sludge on the 18" x 54" scroll centrifuge.



SYMBOL	◐	●	○	⬡	□
Feed.............gpm	15	15	15	15	20
Feed............% SS	2.14	2.14	2.14	2.14	2.14
Bowl Speed.......rpm	1558	1558	1558	1558	1558
Acceleration.....g's	650	650	650	650	650
Rel. Scroll......rpm	2.0	2.7	2.7	3.5	4.9
Pool Depth........in	3.94	3.94	3.94	3.94	3.94
Alum.........lb/ton	100	100	0	0	0

Figure 63. Cake solids vs. polymer dosage for dewatering digested WAS[1] on the 18" x 54" scroll centrifuge.

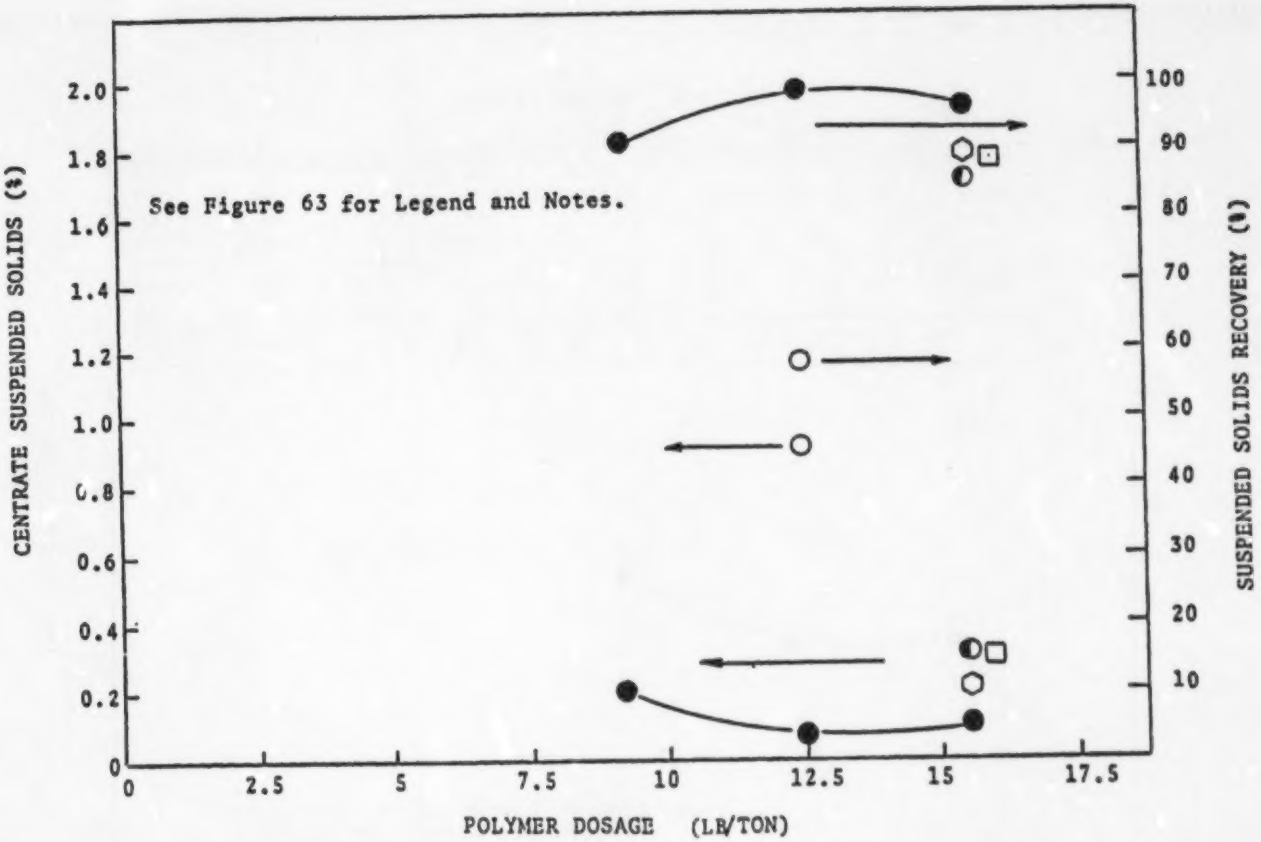
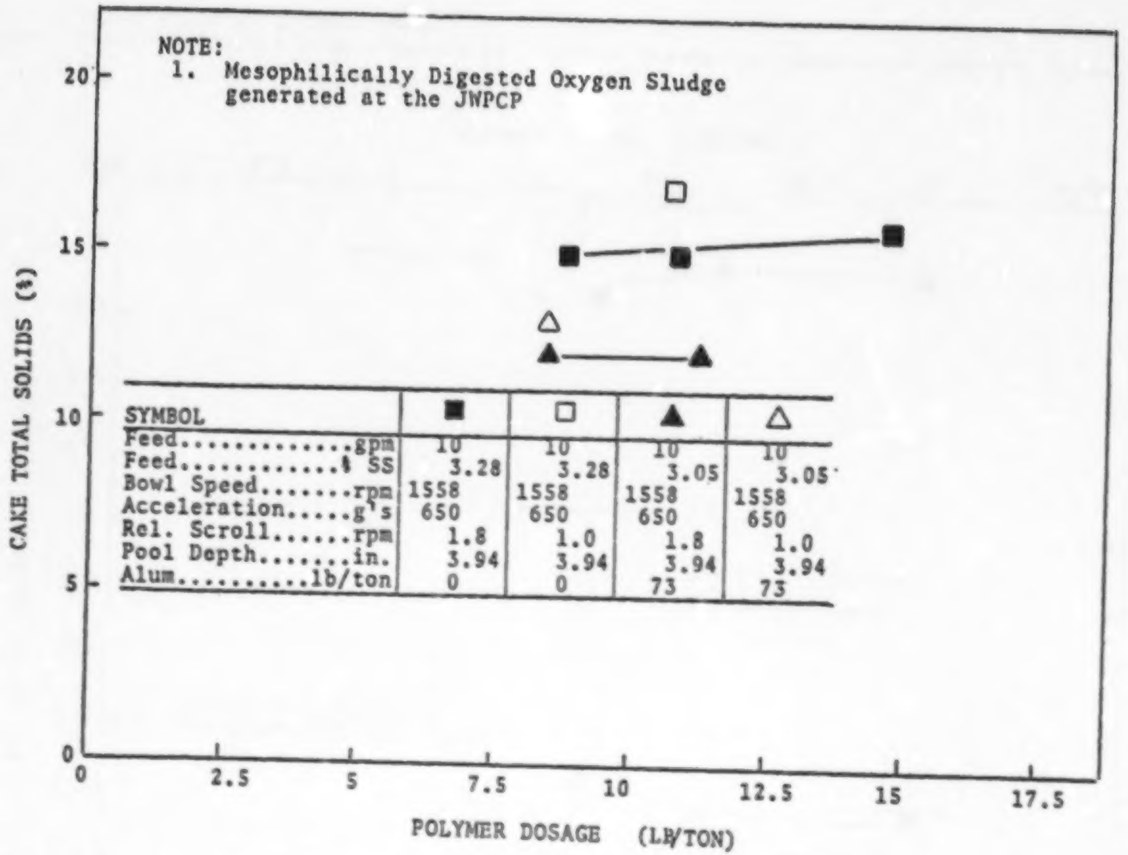

See Figure 63 for Legend and Notes.

Figure 64. Centrate quality & SS recovery vs. polymer dosage for dewatering digested WAS on the 18" x 54" scroll centrifuge.



SYMBOL	■	□	▲	△
Feed..............gpm	10	10	10	10
Feed.............% SS	3.28	3.28	3.05	3.05
Bowl Speed.......rpm	1558	1558	1558	1558
Acceleration.....g's	650	650	650	650
Rel. Scroll......rpm	1.8	1.0	1.8	1.0
Pool Depth.......in.	3.94	3.94	3.94	3.94
Alum..........lb/ton	0	0	73	73

Figure 65. Cake solids vs. polymer dosage for dewatering digested oxygen sludge[1] on the 18" x 54" scroll centrifuge.

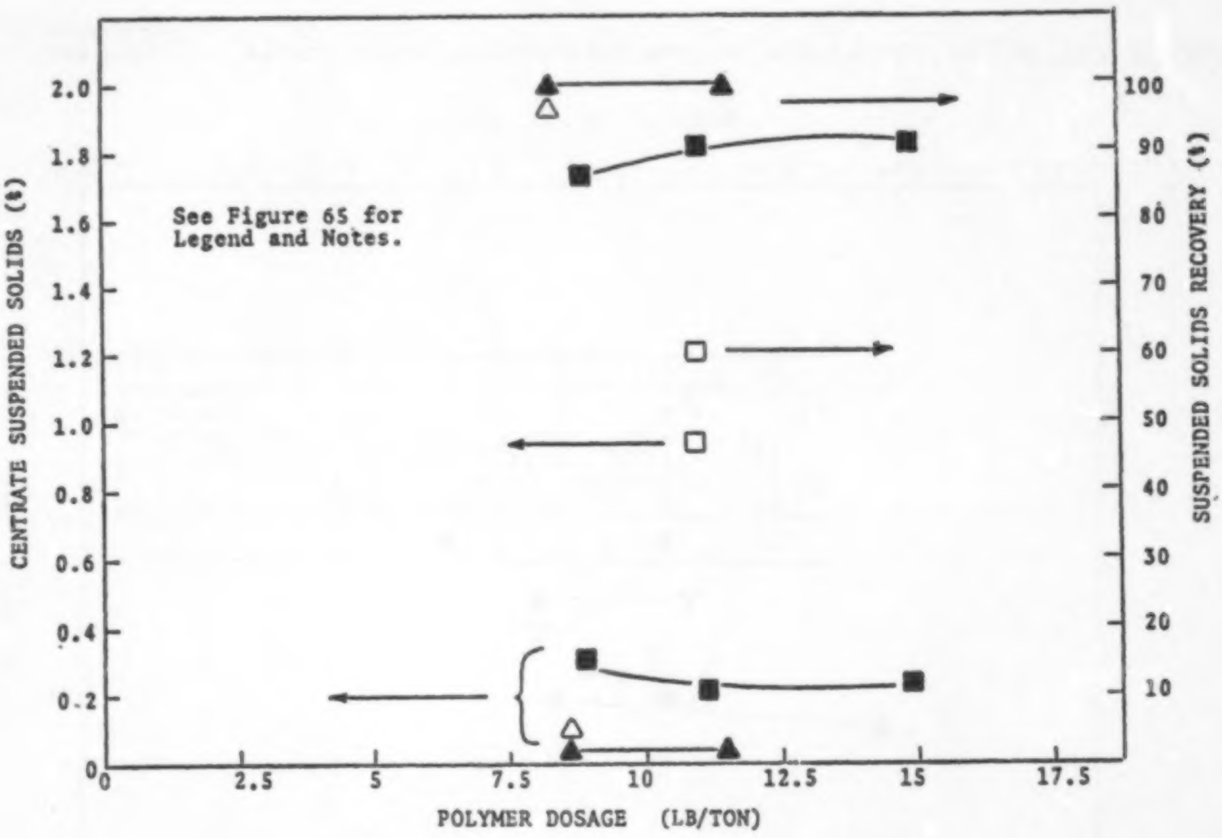
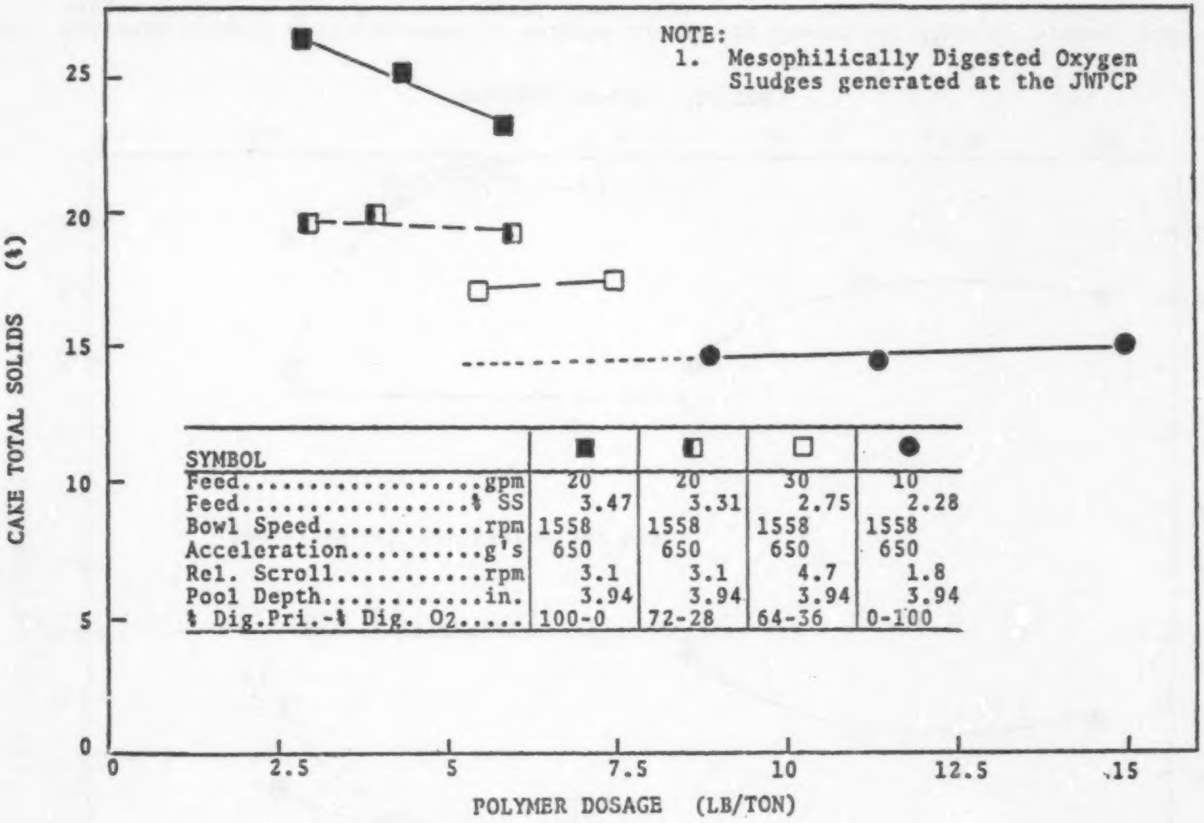
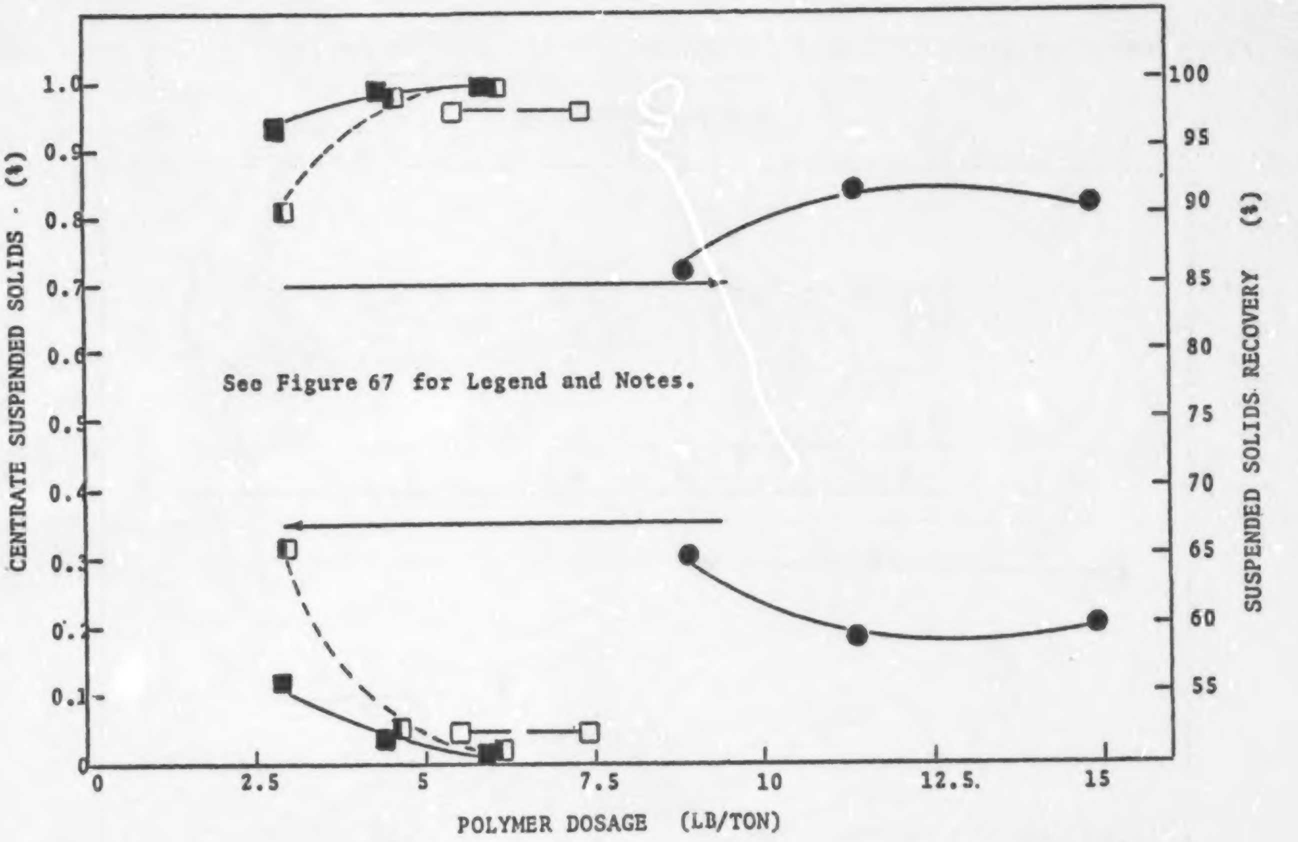


Figure 66. Centrate quality & SS recovery vs. polymer dosage for dewatering digested oxygen sludge on the 18" x 54" scroll centrifuge.



SYMBOL	■	◧	□	●
Feed..................gpm	20	20	30	10
Feed..................% SS	3.47	3.31	2.75	2.28
Bowl Speed............rpm	1558	1558	1558	1558
Acceleration..........g's	650	650	650	650
Rel. Scroll...........rpm	3.1	3.1	4.7	1.8
Pool Depth............in.	3.94	3.94	3.94	3.94
% Dig.Pri.-% Dig. O_2.....	100-0	72-28	64-36	0-100

Figure 67. Cake solids vs. polymer dosage for dewatering digested primary plus digested oxygen sludge[1] on the 18" x 54" scroll centrifuge.



See Figure 67 for Legend and Notes.

Figure 68. Centrate quality & SS recovery vs. polymer dosage for dewatering digested primary plus digested oxygen sludge on the 18" x 54" scroll centrifuge.

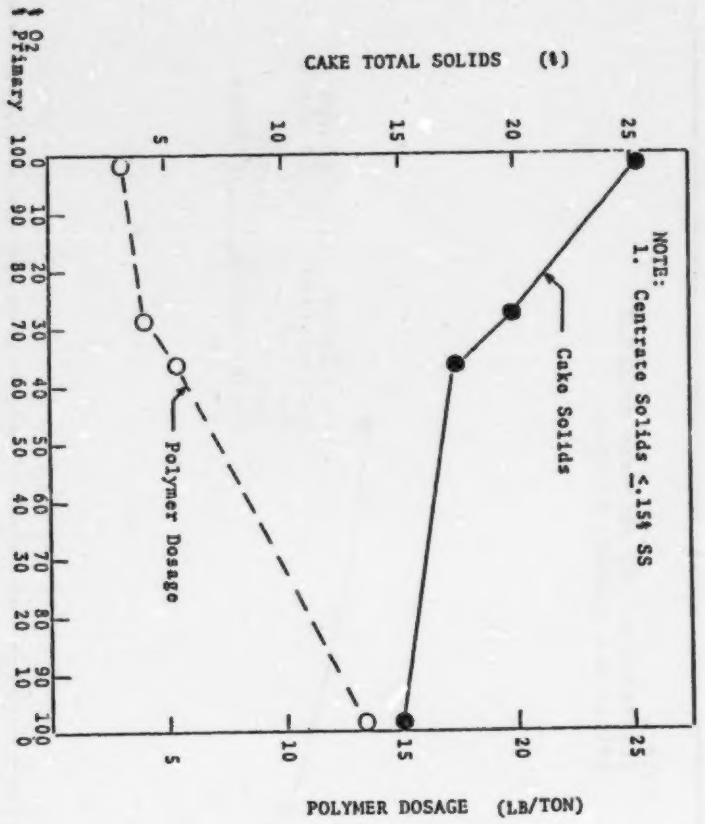


Figure 69. Cake solids and polymer dosage vs. sludge fraction for dewatering digested combined primary and oxygen sludge on the 18" x 54" scroll centrifuge.

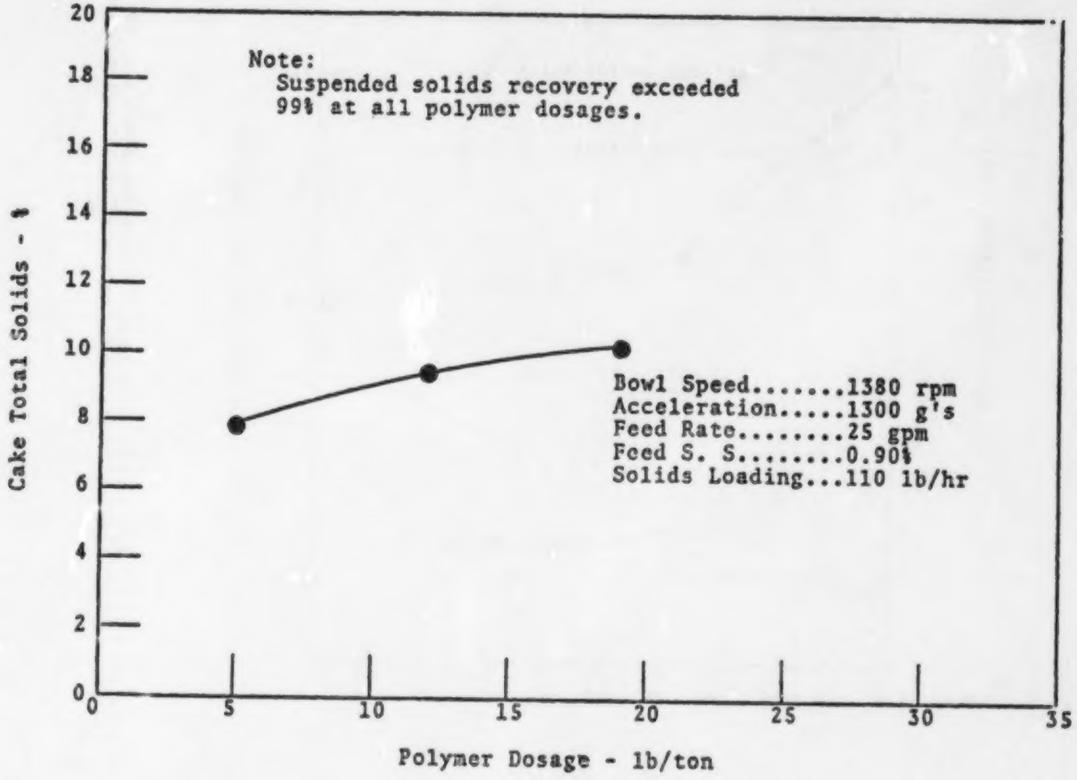


Figure 70. Cake solids vs. polymer dosage for dewatering aerobically digested waste activated sludge on the 48" basket centrifuge.

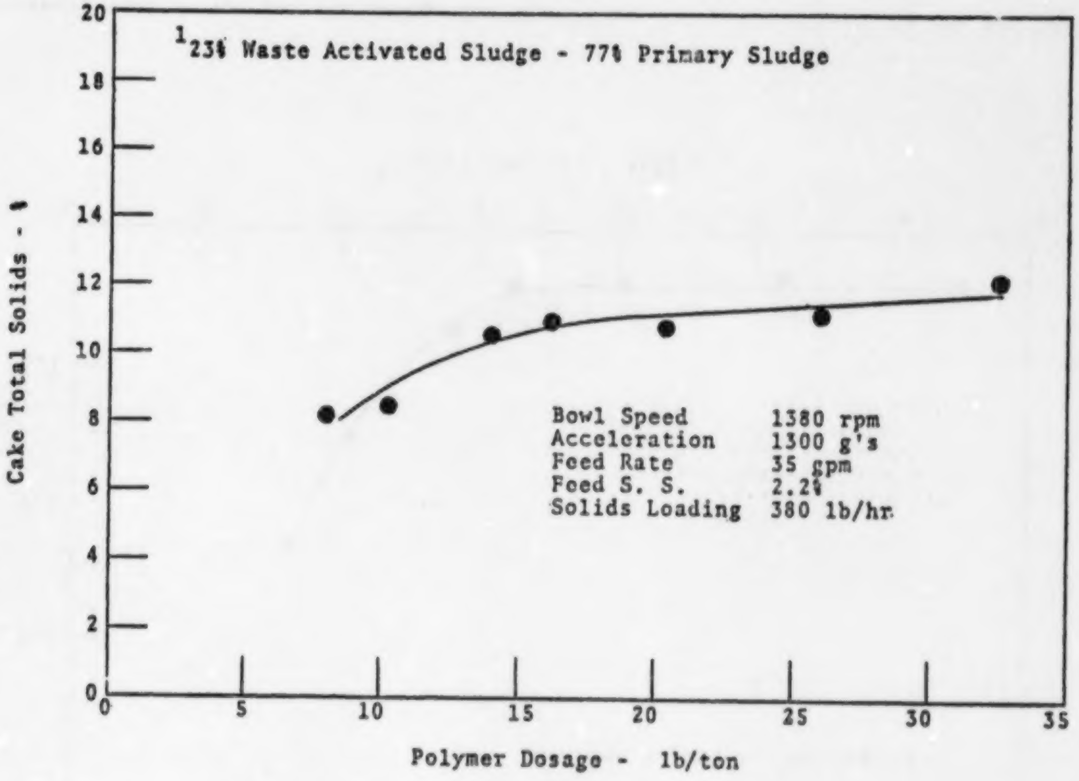
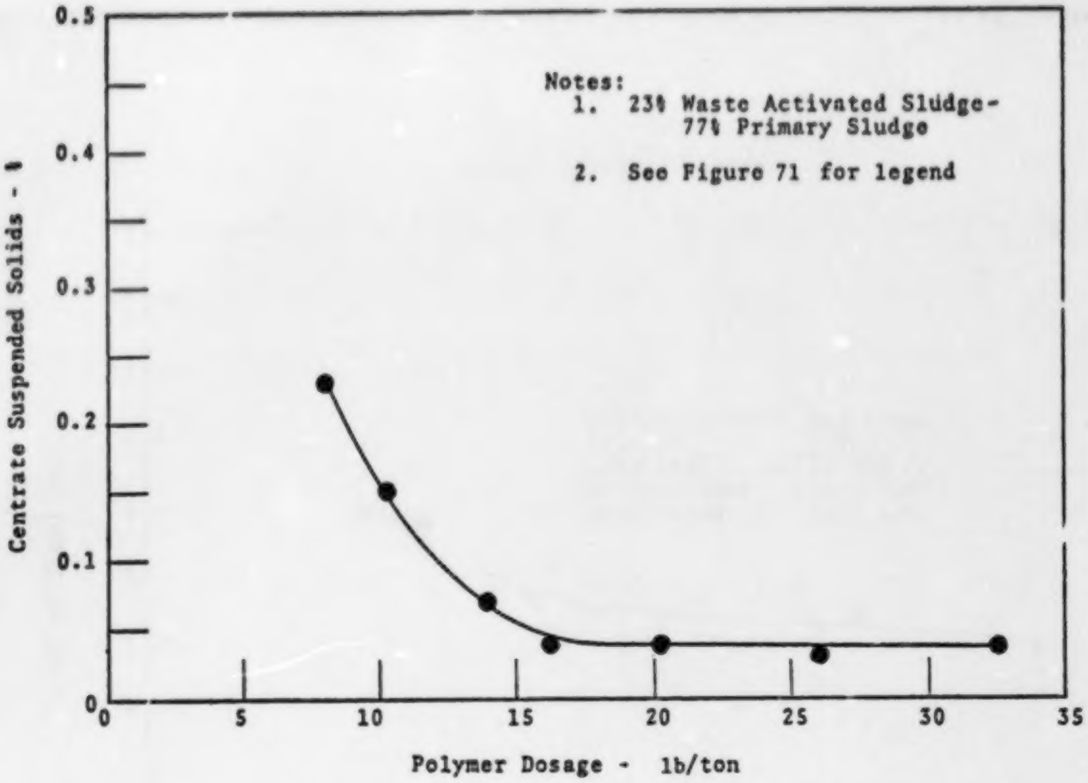


Figure 71. Cake solids vs. polymer dosage for dewatering a digested blend[1] on the 48" basket centrifuge.



Notes:
1. 23% Waste Activated Sludge-
77% Primary Sludge
2. See Figure 71 for legend

Figure 72. Centrate quality vs. polymer dosage for dewatering a digested blend[1] on the 48" basket centrifuge.

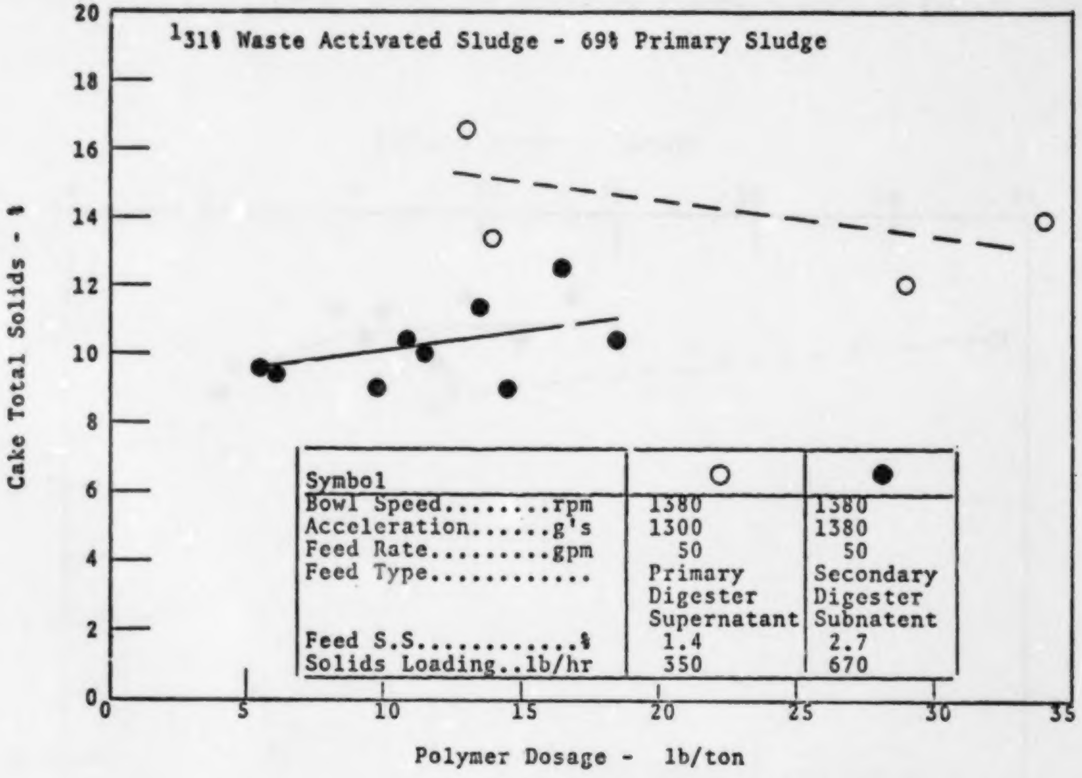


Figure 73. Cake solids vs. polymer dosage for dewatering a digested blend[1] on the 48" basket centrifuge.

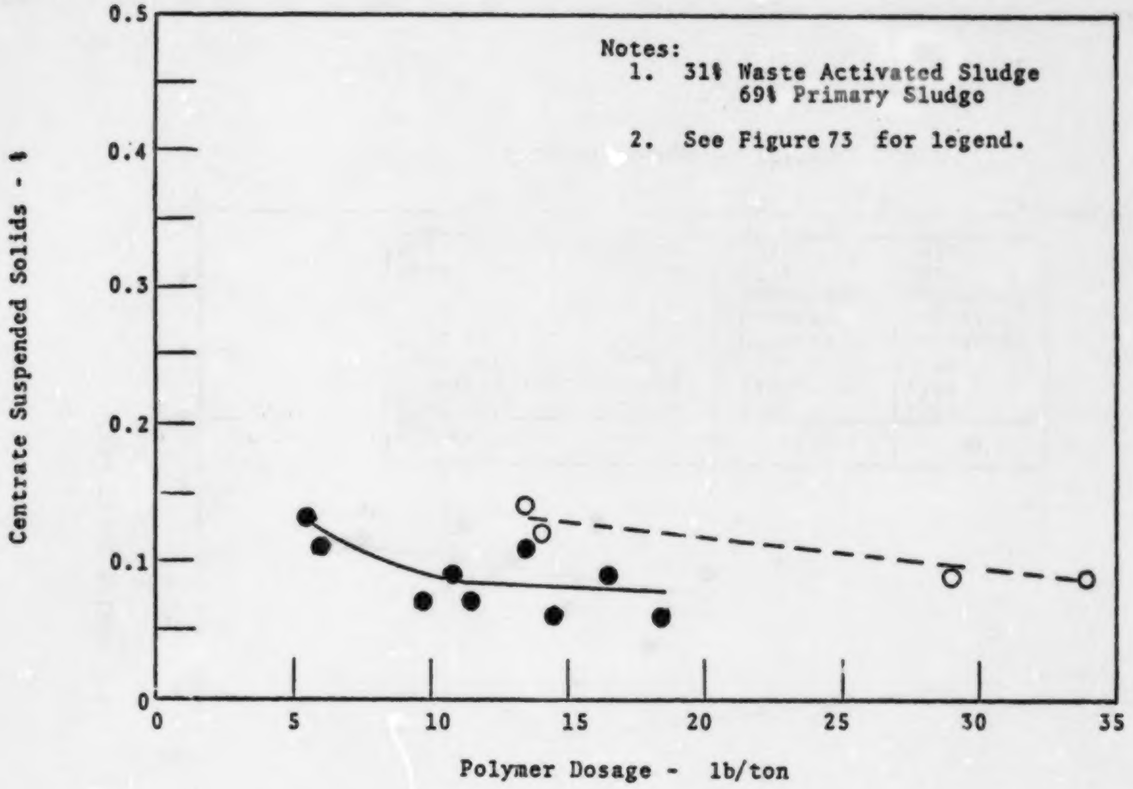


Figure 74. Centrate quality vs. polymer dosage for dewatering a digested blend on the 48" basket centrifuge.

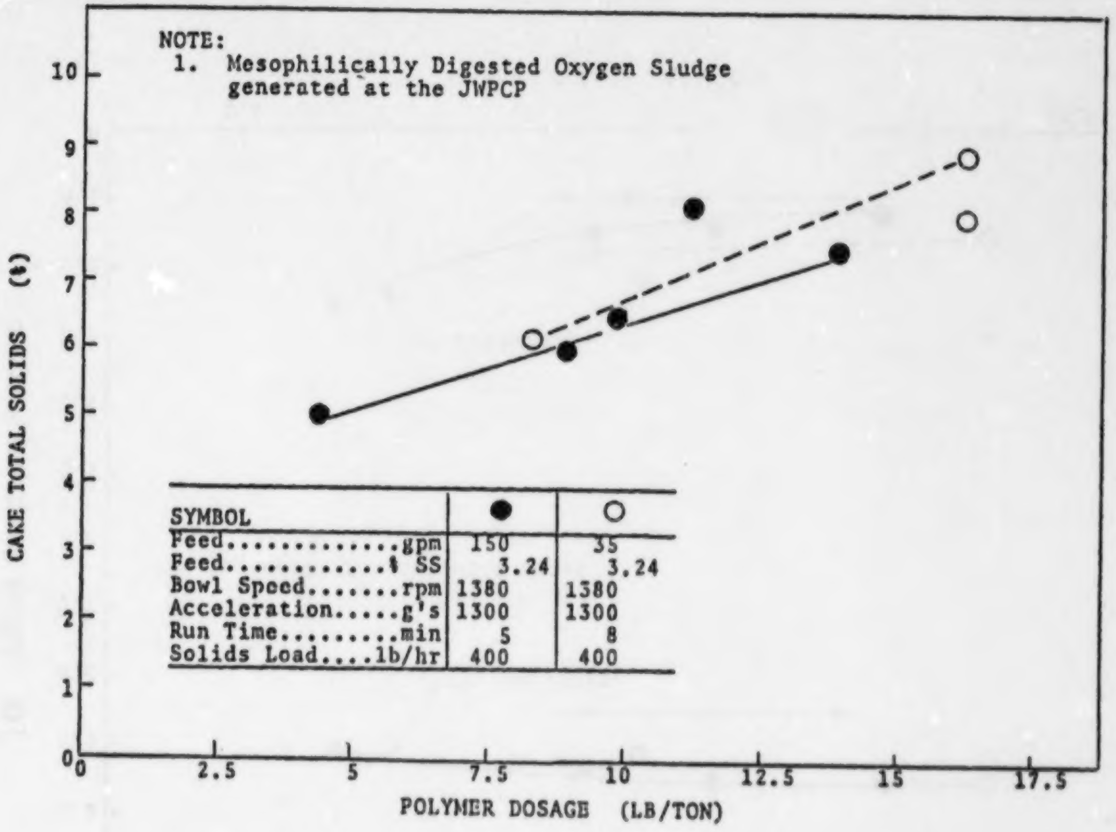


Figure 75. Cake solids vs. polymer dosage for dewatering digested oxygen sludge on the 48" basket centrifuge.

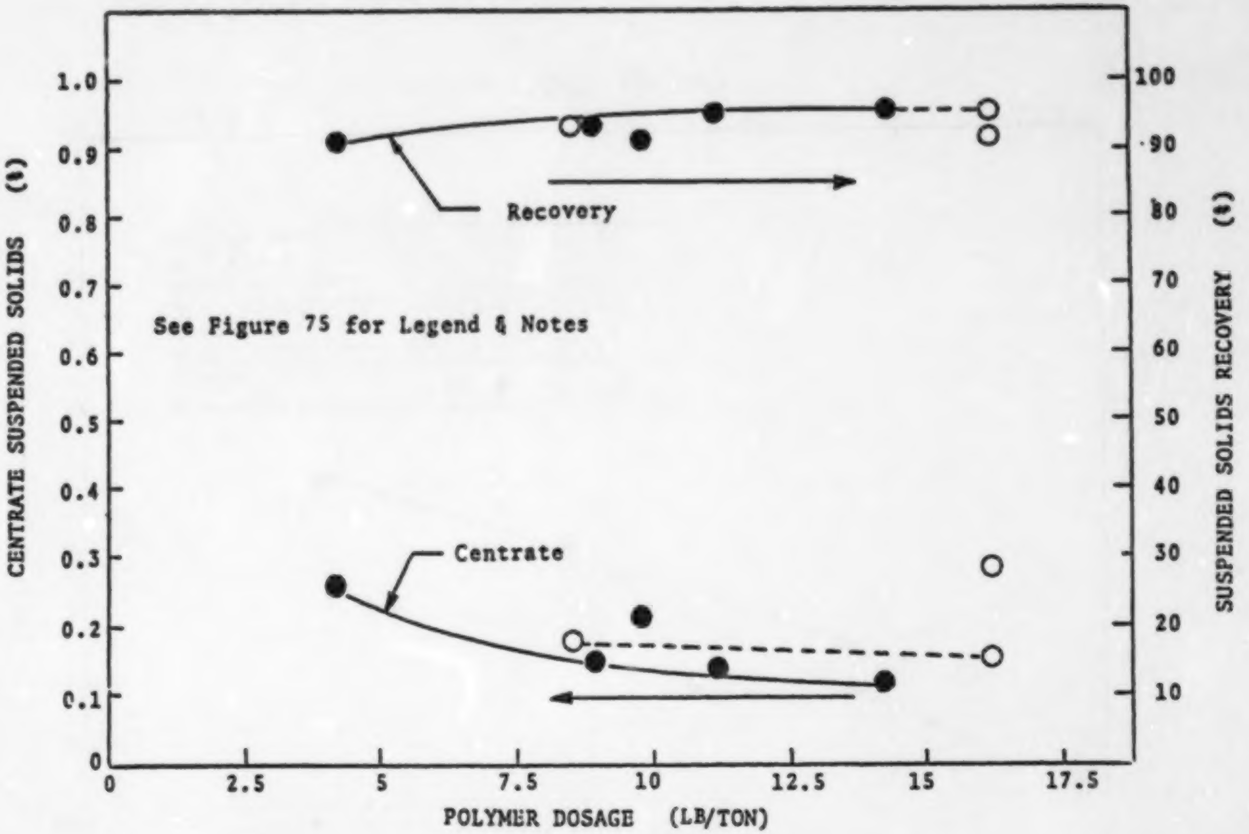


Figure 76. Centrate quality & SS recovery vs. polymer dosage for dewatering digested oxygen sludge on the 48" basket centrifuge.

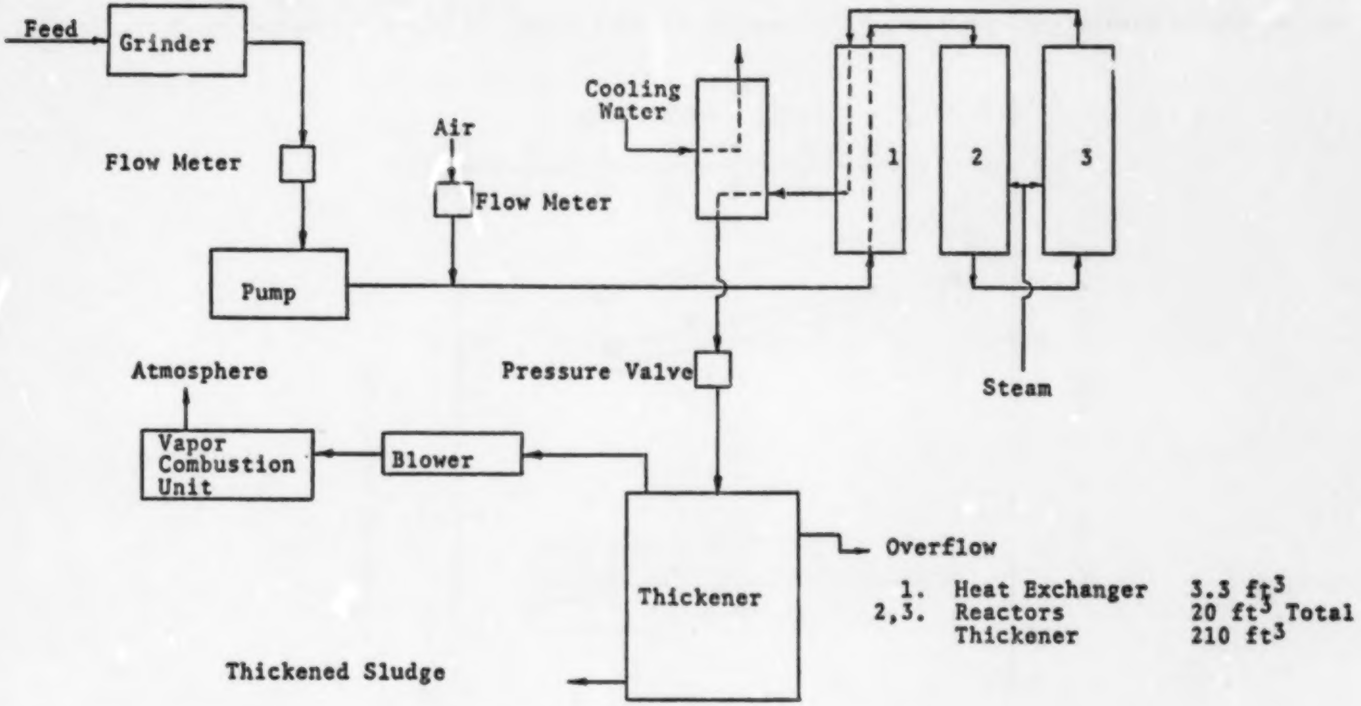

Feed
Grinder
Flow Meter
Air
Flow Meter
Pump
Cooling Water
1
2
3
Steam
Pressure Valve
Atmosphere
Vapor Combustion Unit
Blower
Thickener
Overflow
Thickened Sludge
1. Heat Exchanger 3.3 ft³
2,3. Reactors 20 ft³ Total
Thickener 210 ft³

Figure 77. Thermal conditioning schematic

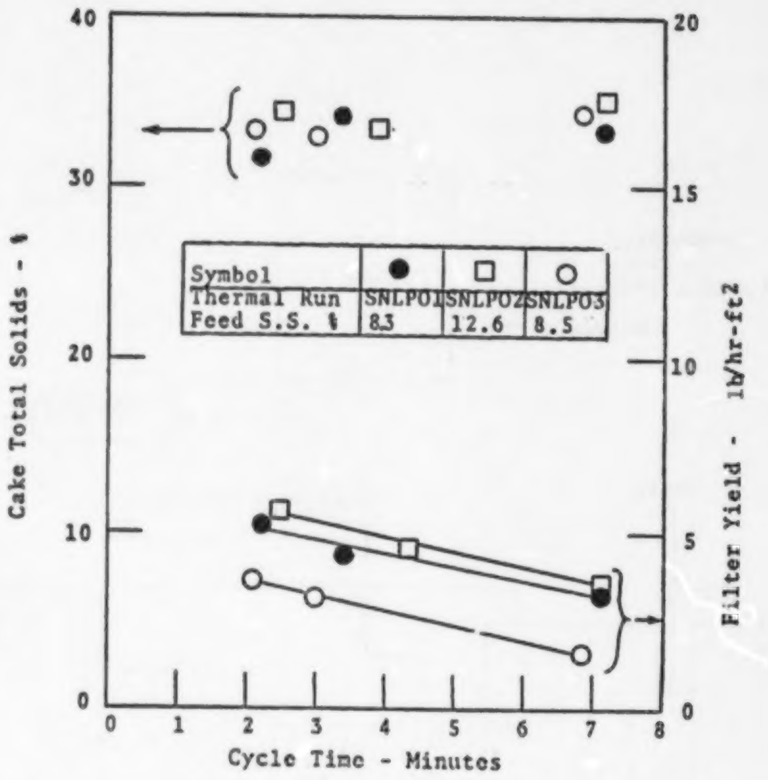


Figure 78. Cake solids and yield vs. cycle time for dewatering LPO thermal conditioned sludge on the 3' x 1' rotary drum vacuum filter.

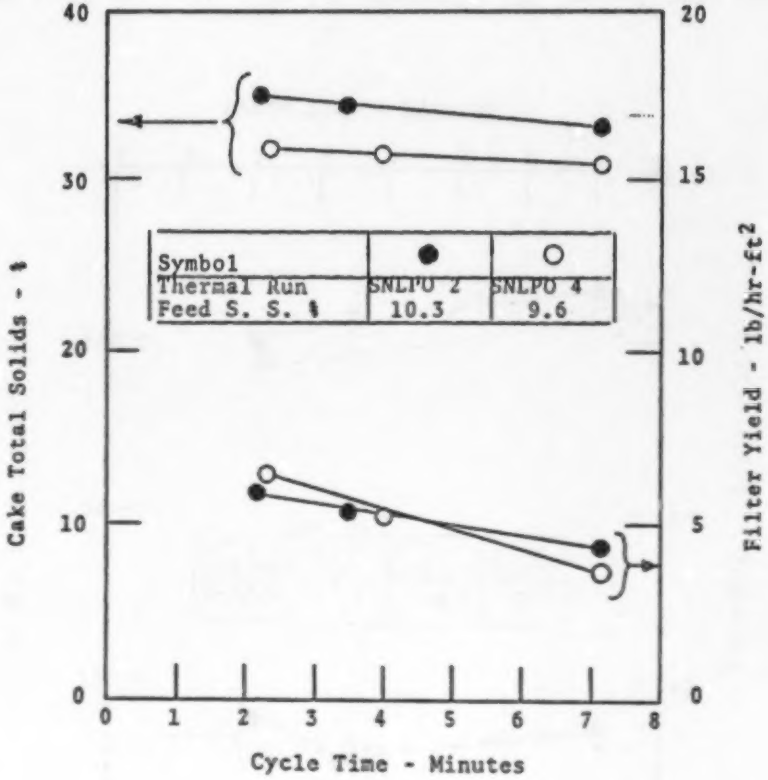


Figure 79. Cake solids and yield vs. cycle time for dewatering LPO thermal conditioned sludge on the 3' x 1' rotary drum vacuum filter.

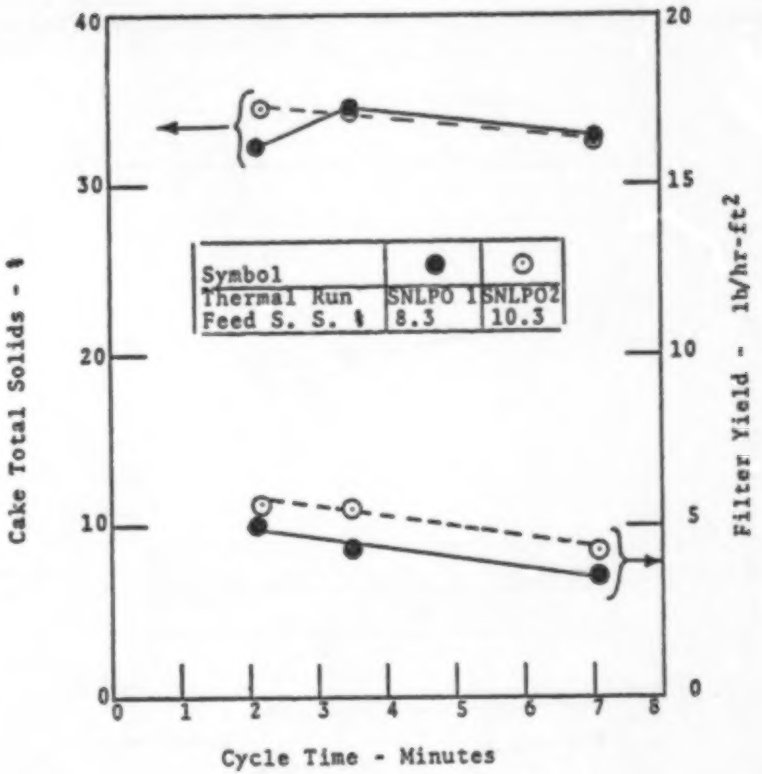


Figure 80. Cake solids and yield vs. cycle time for dewatering LPO thermal conditioned sludge on the 3' x 1' rotary drum vacuum filter.

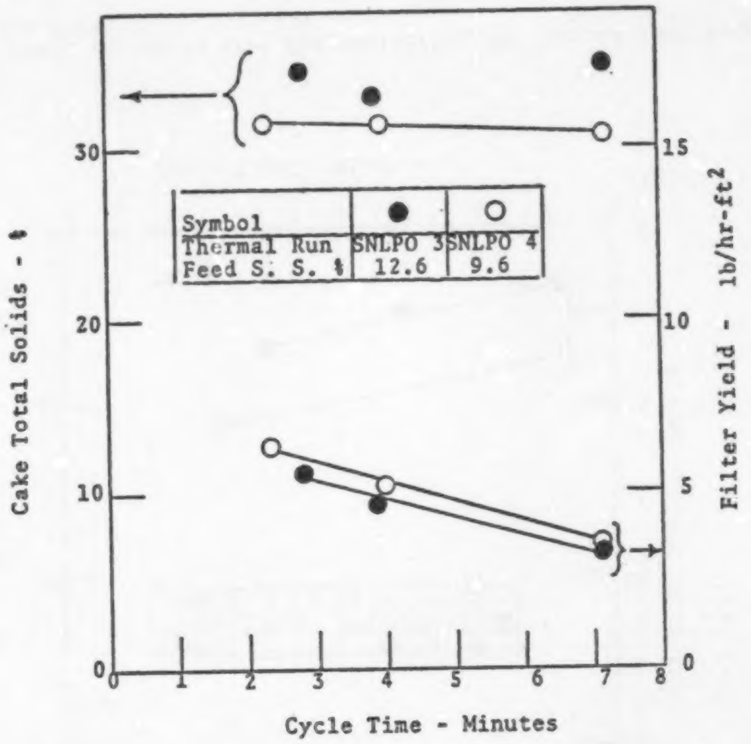


Figure 81. Cake solids and filter yield vs. cycle time for dewatering LPO thermal conditioned sludge on the 3' x 1' rotary drum vacuum filter.

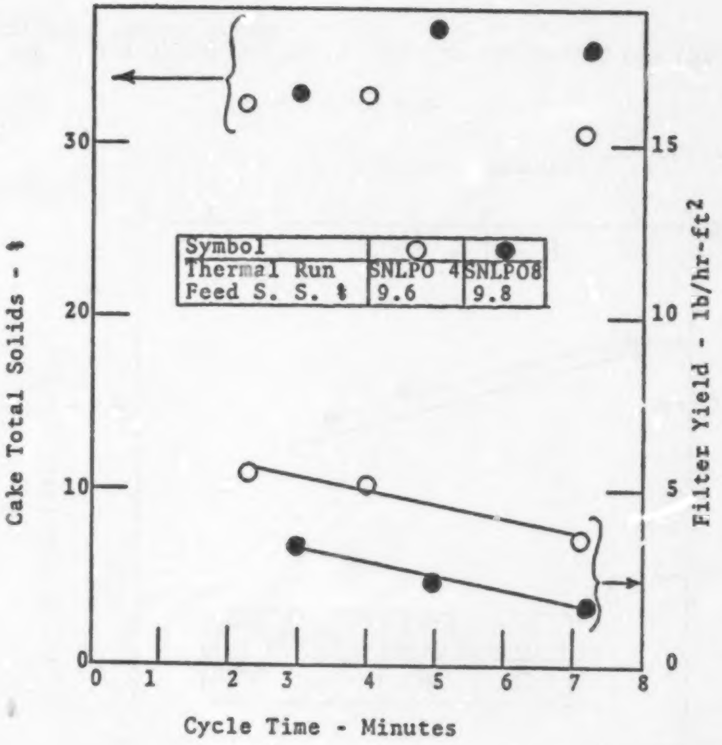


Figure 82. Cake solids and yield vs. cycle time for dewatering LPO thermal conditioned sludge on the 3' x 1' rotary drum vacuum filter.

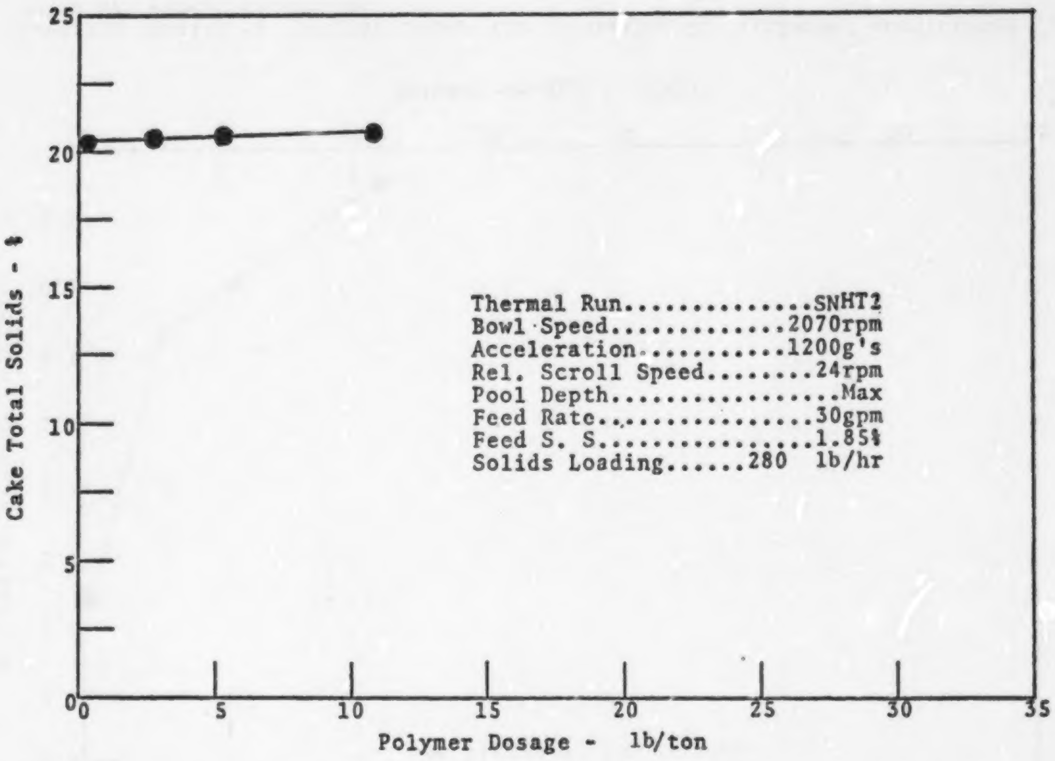


Figure 83. Cake solids vs. polymer dosage for dewatering H. T. thermal conditioned sludge on the 20" x 62" scroll centrifuge.

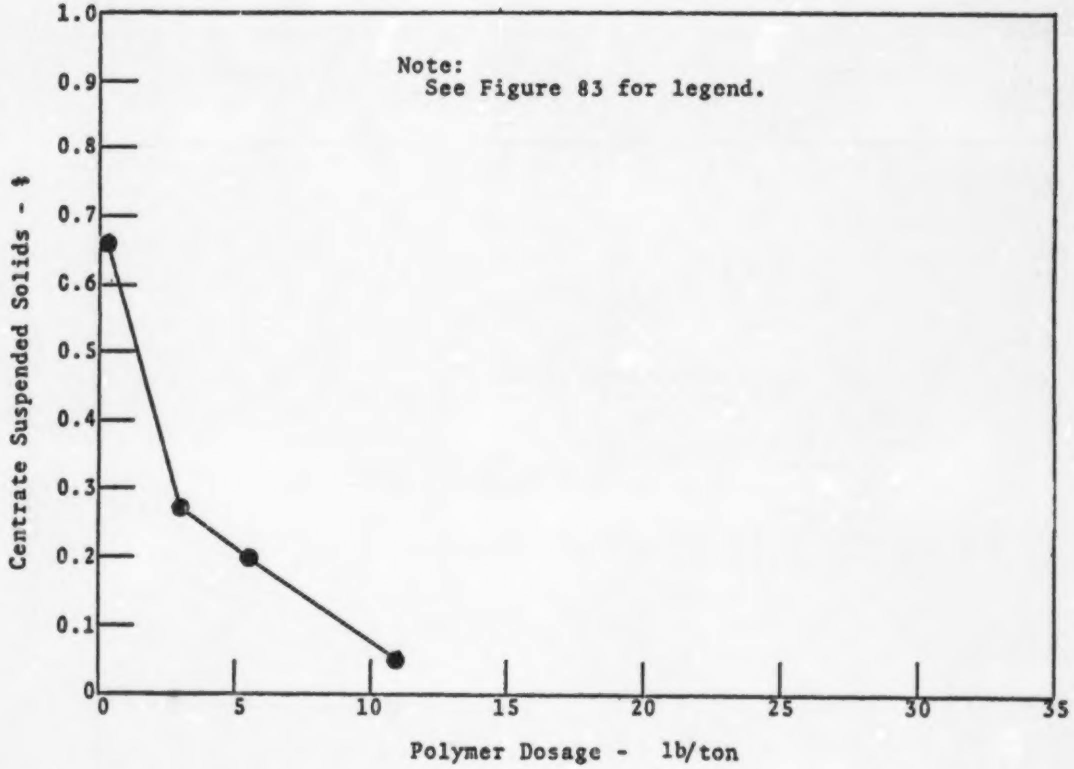


Figure 84. Centrate quality vs. polymer dosage for dewatering H. T. thermal conditioned sludge on the 20" x 62" scroll centrifuge.

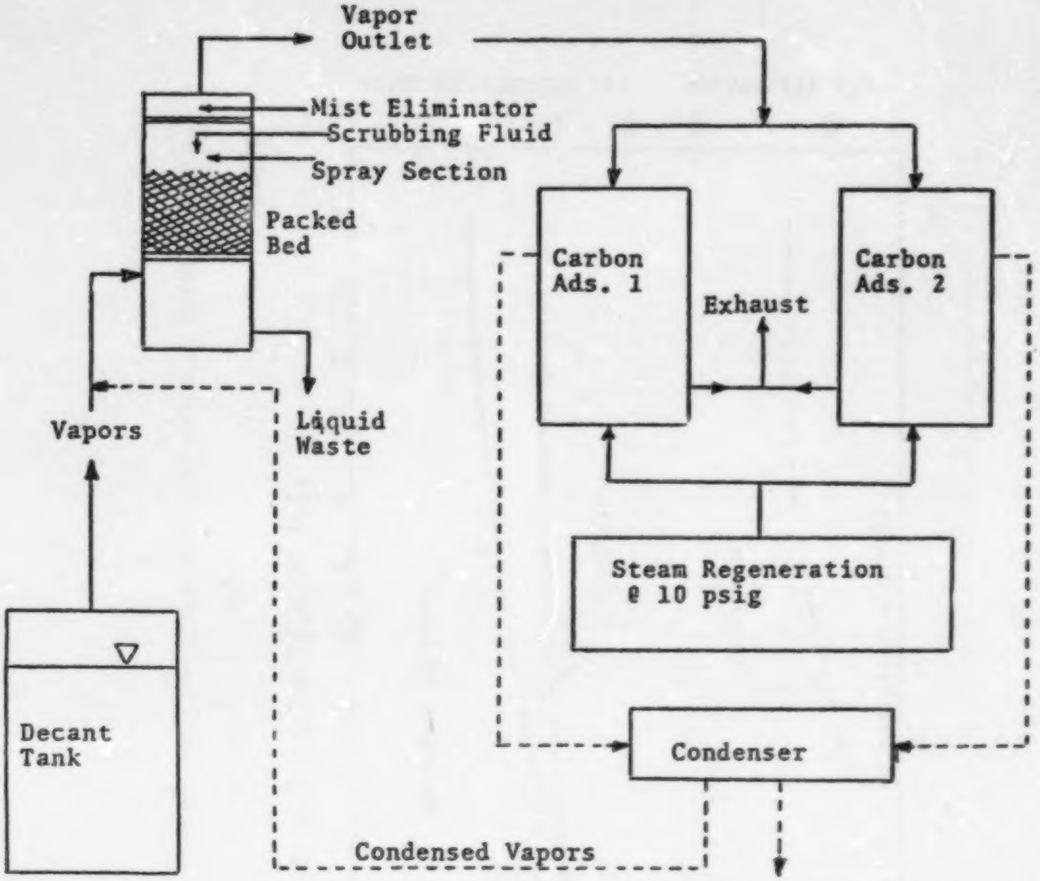

Vapor Outlet
Mist Eliminator
Scrubbing Fluid
Spray Section
Packed Bed
Vapors
Liquid Waste
Decant Tank
Carbon Ads. 1
Carbon Ads. 2
Exhaust
Steam Regeneration @ 10 psig
Condenser
Condensed Vapors

Figure 85. Odor control schematic

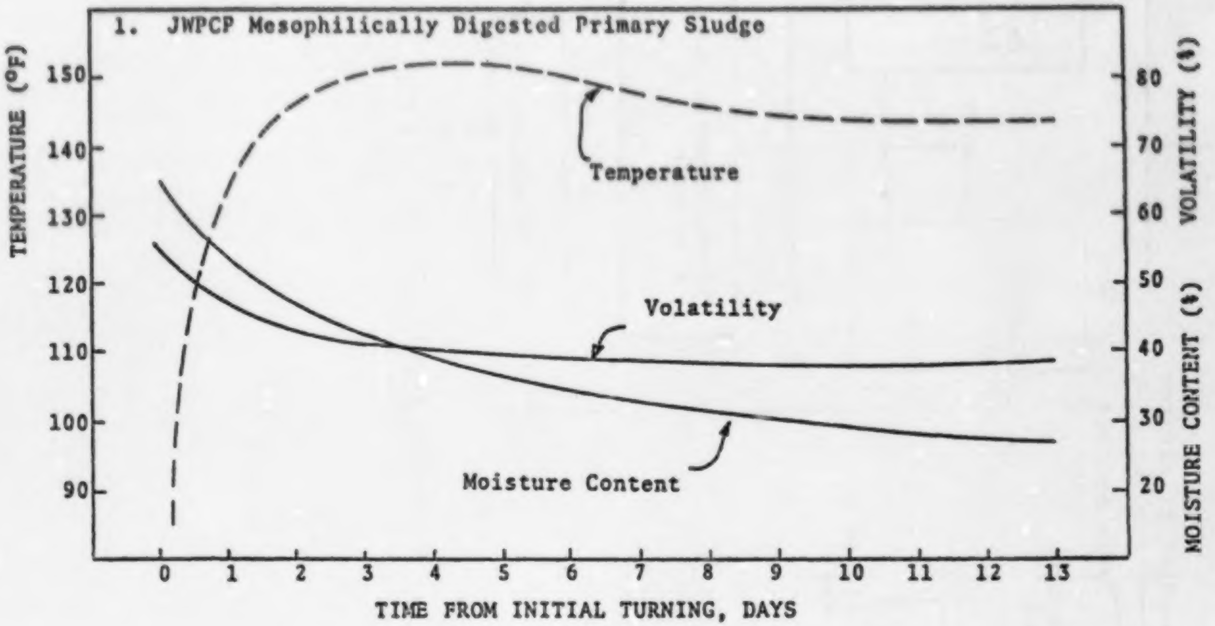


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Figure 86. Compost parameters vs. drying time for composting dewatered digested primary sludge[1].

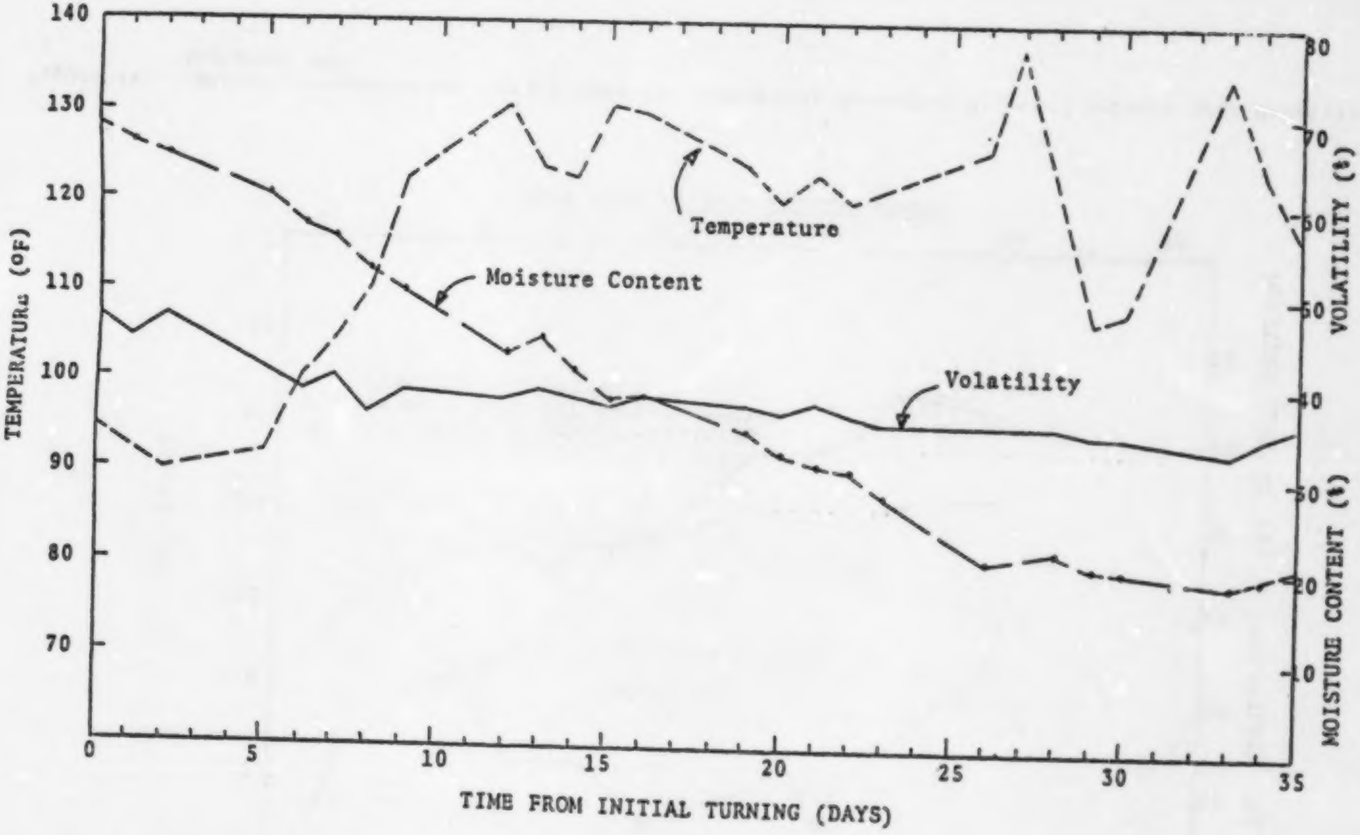

Figure 87. Compost parameters vs. drying time for composting dewatered digested WAS[1].

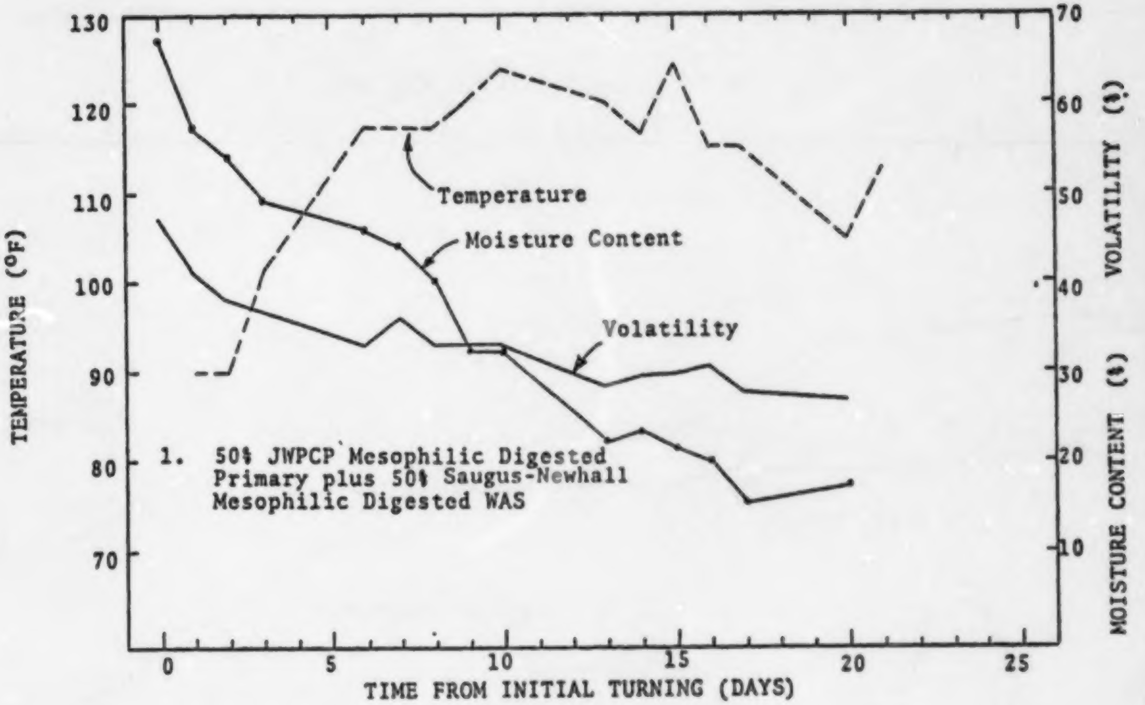

Figure 88. Compost parameters vs. drying time for composting dewatered digested primary plus dewatered digested WAS[1].

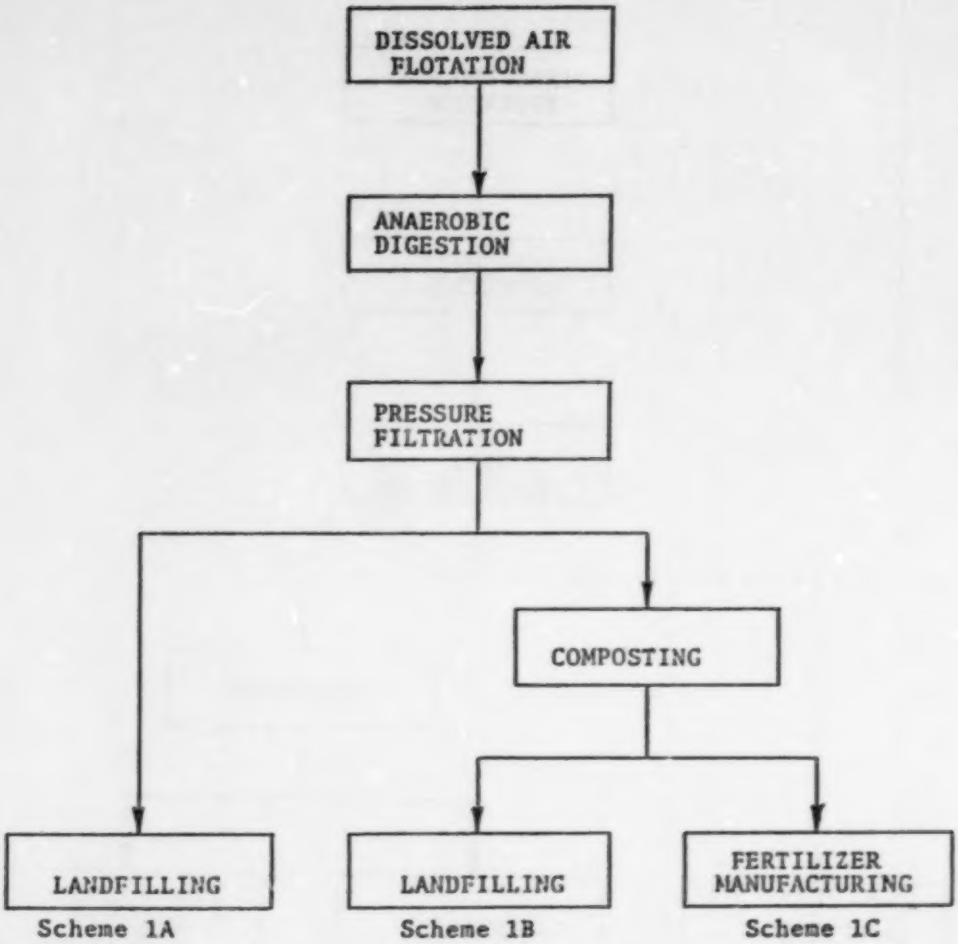
DISSOLVED AIR FLOTATION
ANAEROBIC DIGESTION
PRESSURE FILTRATION
COMPOSTING
LANDFILLING
Scheme 1A
LANDFILLING
Scheme 1B
FERTILIZER MANUFACTURING
Scheme 1C

Figure 89. Economic Analysis: Sludge handling scheme 1.

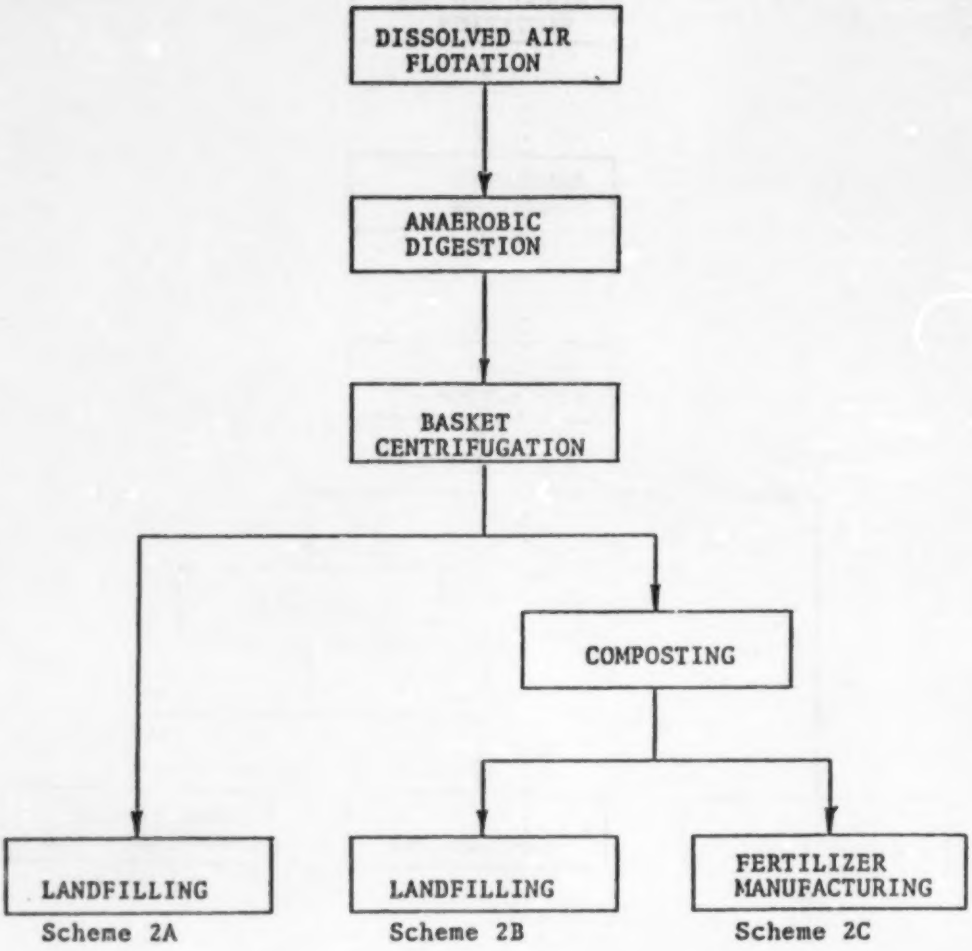
DISSOLVED AIR FLOTATION
ANAEROBIC DIGESTION
BASKET CENTRIFUGATION
COMPOSTING
LANDFILLING
Scheme 2A
LANDFILLING
Scheme 2B
FERTILIZER MANUFACTURING
Scheme 2C

Figure 90. Economic Analysis: Sludge handling scheme 2.

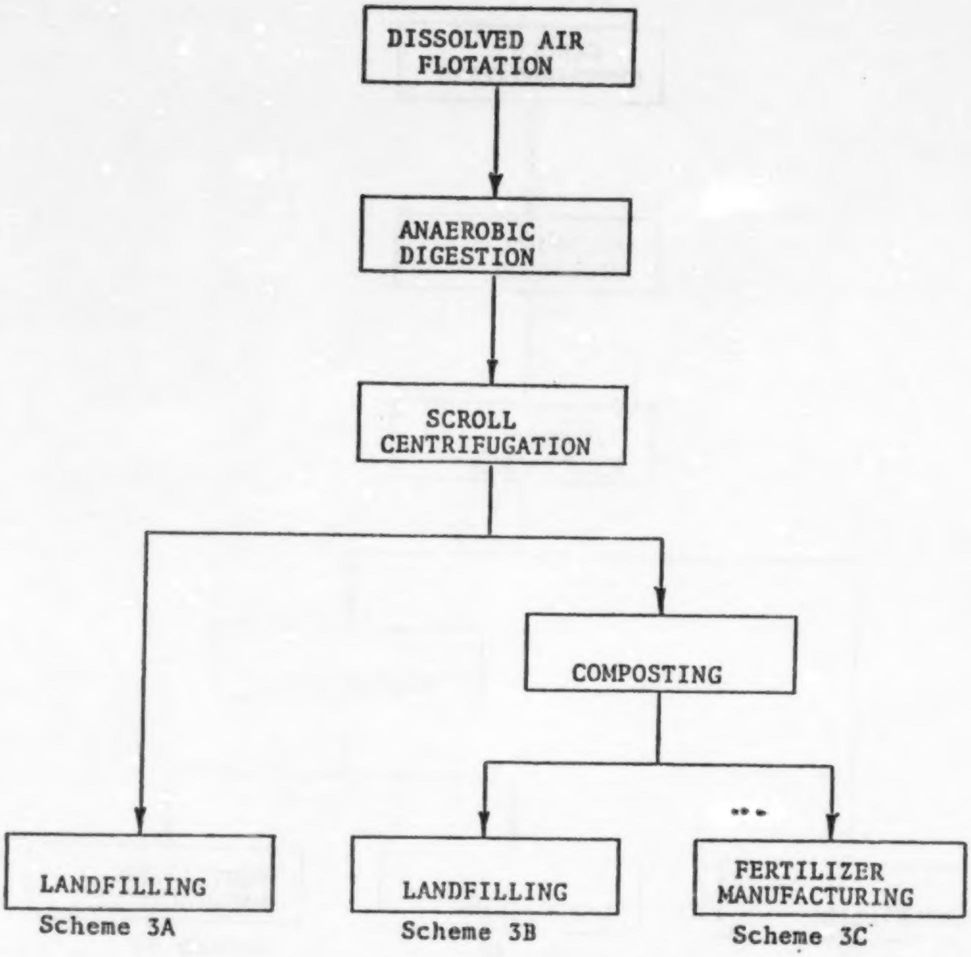
DISSOLVED AIR FLOTATION
ANAEROBIC DIGESTION
SCROLL CENTRIFUGATION
COMPOSTING
LANDFILLING
Scheme 3A
LANDFILLING
Scheme 3B
FERTILIZER MANUFACTURING
Scheme 3C

Figure 91. Economic Analysis: Sludge handling scheme 3.

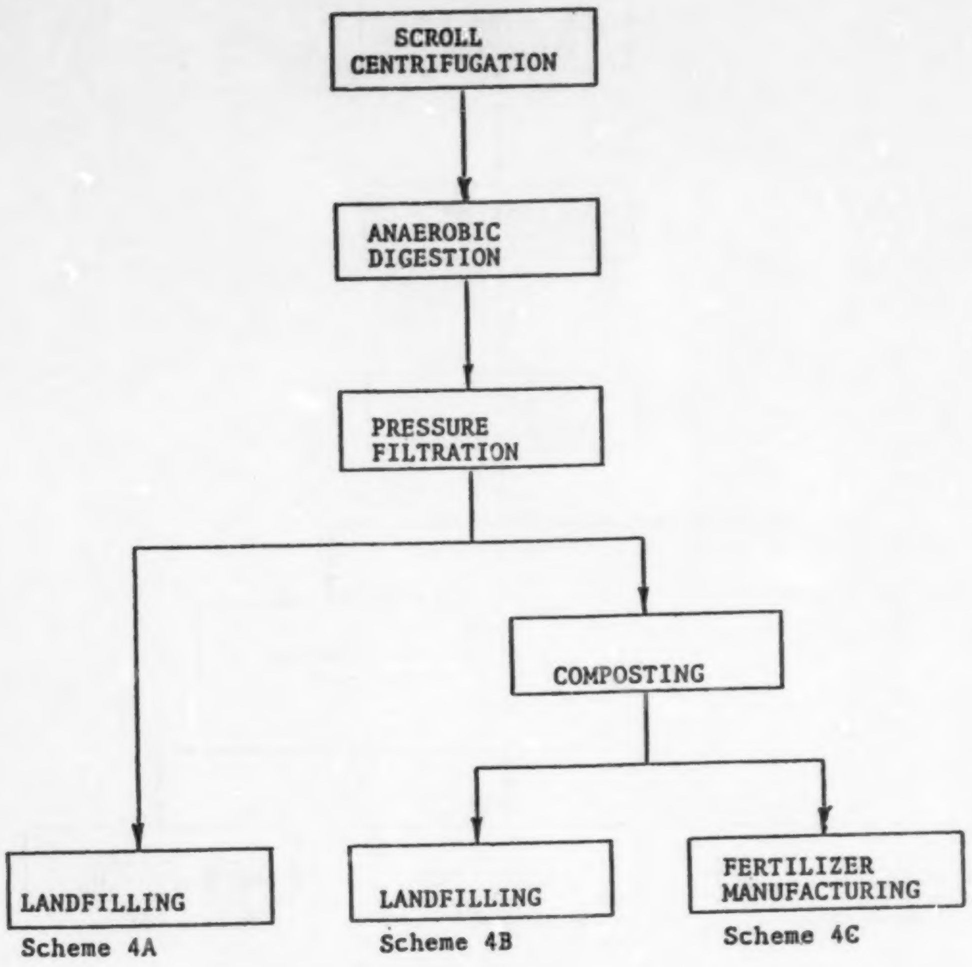
SCROLL CENTRIFUGATION
ANAEROBIC DIGESTION
PRESSURE FILTRATION
COMPOSTING
LANDFILLING
Scheme 4A
LANDFILLING
Scheme 4B
FERTILIZER MANUFACTURING
Scheme 4C

Figure 92. Economic Analysis: Sludge handling scheme 4.

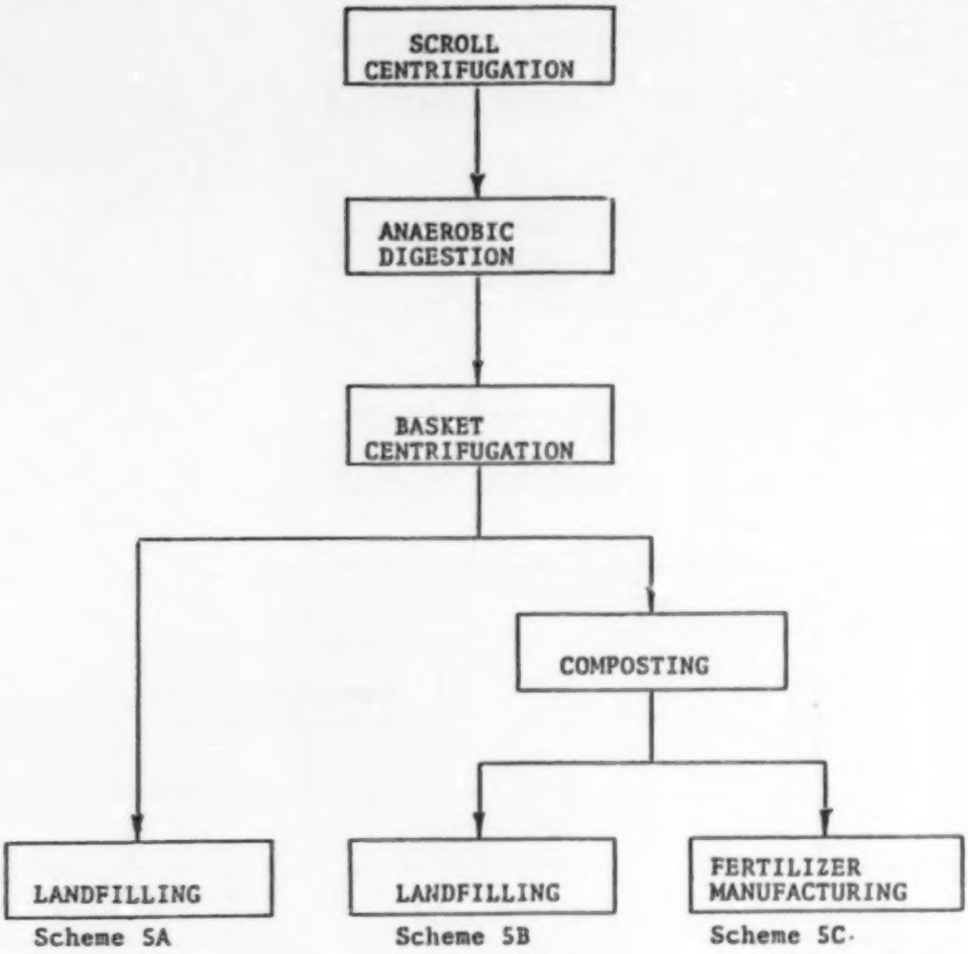
SCROLL CENTRIFUGATION
ANAEROBIC DIGESTION
BASKET CENTRIFUGATION
COMPOSTING
LANDFILLING
Scheme 5A
LANDFILLING
Scheme 5B
FERTILIZER MANUFACTURING
Scheme 5C

Figure 93. Economic Analysis: Sludge handling scheme 5.

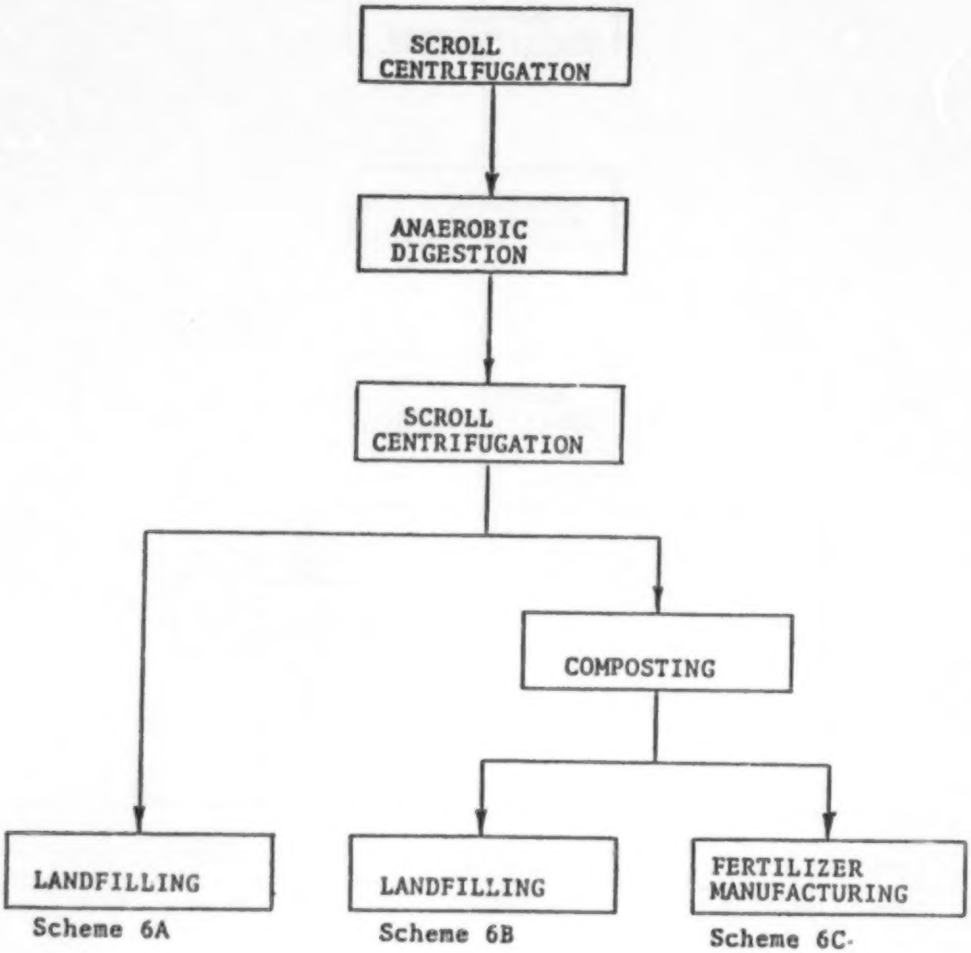
SCROLL CENTRIFUGATION
ANAEROBIC DIGESTION
SCROLL CENTRIFUGATION
COMPOSTING
LANDFILLING
Scheme 6A
LANDFILLING
Scheme 6B
FERTILIZER MANUFACTURING
Scheme 6C

Figure 94. Economic Analysis: Sludge handling scheme 6.

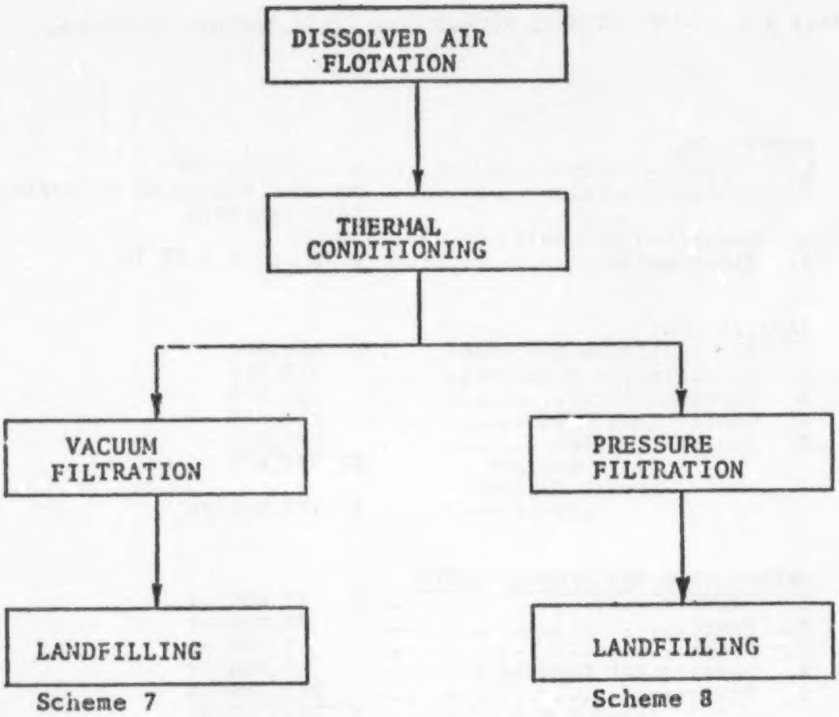
DISSOLVED AIR FLOTATION
THERMAL CONDITIONING
VACUUM FILTRATION
PRESSURE FILTRATION
LANDFILLING
LANDFILLING
Scheme 7
Scheme 8

Figure 95. Economic analysis: Sludge handling schemes 7 and 8.

APPENDIX A

ECONOMIC ANALYSIS CALCULATIONS

TABLE A-1. COST ESTIMATE SUMMARY FOR AIR FLOTATION THICKENING

CONDITIONS

1. Suspended Solids Loading	3 lb SS/hr-ft^2
2. Sludge Conditioning	polymer dosage of 3 lbs/to (318 lbs/day)
3. Underflow Suspended Solids	50 mg/l
4. Float Solids	0.73 mgd @ 3.5% TS

CAPITAL COST

1. Air Flotation-purchased	$ 840,000
2. Installation & Housing	210,000
3. Contingencies	210,000
4. Contractors Profit	110,000
5. Engineering Fee	170,000
Total Capital	$1,540,000
Annual Capital (7%-15 yrs)	$ 170,000/yr

OPERATION & MAINTENANCE COSTS

1. Labor	$ 57,000/yr
2. Power	30,000/yr
3. Water	2,300/yr
4. Maintenance Materials	15,000/yr
5. Polymers	240,000/yr
Total O & M	$ 344,300/yr
Total Annual	$ 514,300/yr
$/ton	$ 13.30/ton

TABLE A-2. COST ESTIMATE SUMMARY FOR SCROLL CENTRIFUGE THICKENING

CONDITIONS	
1. Suspended Solids Loading........	600 lbs/hr - 32" x 100" unit
2. Sludge Conditioning............	polymer dosage of 10 lbs/ton (1060 lbs/day)
3. Centrate Suspended Solids.......	600 mg/l
4. Thickened Sludge................	0.40 MGD @ 6% TS

CAPITAL COST	
1. 15 Scroll Centrifuges-purchased.	$2,550,000
2. Installation & Housing..........	640,000
3. Contingencies...................	640,000
4. Contractors Profit..............	320,000
5. Engineering Fees................	510,000
Total Capital	$4,660,000
Annual Capital (7%-15 yrs)	$ 510,000/yr

OPERATION & MAINTENANCE COSTS	
1. Labor...........................	$ 114,000
2. Power...........................	170,000
3. Water...........................	4,600
4. Maintenance Materials...........	46,000
5. Polymers........................	780,000
Total O & M	$1,114,600/yr
Total Annual	$1,624,600/yr
$/ton	$ 42.00/ton

TABLE A-3. COST ESTIMATE SUMMARY FOR ANAEROBIC DIGESTION

CONDITIONS

1. Detention Time....................20 days
2. Volume.............................1.75 MCF for flotation thickening
   1.07 MCF for centrifuge thickening

	Thickening Process	
OPERATION COST	Flotation	Centrifugation
1. Construction*...............	$ 8,940,000	$5,500,000
2. Engineering Fees............	1,790,000	1,100,000
Total Capital...............	$10,730,000	$6,600,000
Annual Capital (7%-25 yrs)	$ 922,780/yr	$ 568,000/yr
OPERATION & MAINTENANCE COST		
1. Labor.........................	$ 114,000/yr	$ 114,000/yr
2. Power (Mixing)+............	55,700/yr	55,700/yr
3. Water.........................	3,800/yr	2,400/yr
4. Maintenance Materials.....	103,500/yr	67,500/yr
Total O & M	$ 277,000/yr	$ 239,600/yr
Total Annual	$ 1,199,800/yr	$ 807,600/yr
$/ton	$ 31.00/ton	$ 20.90/ton

*Includes contingencies, contractors profit, control building, heating, gas and electrical equipment.

+Includes credit for Digester Gas.

TABLE A-4. COST ESTIMATE SUMMARY FOR BASKET CENTRIFUGATION DEWATERING OF BLENDED DIGESTED SLUDGES

CONDITIONS	
1. Sludge Quantity	116 tons/day
2. Sludge Blend	70% WAS - 30% Primary
3. Solids Loading	550 lbs/hr - 48" unit
4. Sludge Conditioning	Polymer Dosage of 11 lbs/ton (1276 lbs/day)
5. Centrate Suspended Solids	1500 mg/l
6. Total Cake Solids	11% TS

CAPITAL COST	
1. 20 Basket Centrifuges-purchase	$1,600,000
2. Installation & Housing	400,000
3. Contingencies	400,000
4. Contractors Profit	200,000
5. Engineering Fee	320,000
Total Capital	$2,920,000
Annual Capital(7%-15 yrs)	320,000/yr

OPERATION & MAINTENANCE	
1. Labor	$ 114,000/yr
2. Power	150,000/yr
3. Water	3,600/yr
4. Maintenance Materials	29,200/yr
5. Polymer*	860,000/yr
Total O & M	$1,156,800/yr
Total Annual	$1,476,800/yr
$/ton	34.90/ton

*Includes credit for sludge taken from Primary Sludge station.

TABLE A-5. COST ESTIMATE SUMMARY FOR SCROLL CENTRIFUGE DEWATERING OF BLENDED DIGESTED SLUDGES

CONDITIONS	
1. Sludge Quantity	116 tons/day
2. Sludge Blend	70% WAS - 30% Primary
3. Solids Loading	1500 lbs/hr - 32" x 100" uni
4. Sludge Conditioning	polymer dosage of 16 lbs/ton (1856 lbs/day)
5. Centrate Suspended Solids	1500 mg/l
6. Total Cake Solids	15% TS

CAPITAL COST	
1. 7-Scroll Centrifuges-purchase	$1,190,000
2. Installation & Housing	297,500
3. Contingencies	297,500
4. Contractors Profit	148,800
5. Engineering Fee	238,000
Total Capital	$2,171,800
Annual Capital (7% - 15 yrs)	$ 239,000

OPERATION & MAINTENANCE	
1. Labor	$ 76,000/yr
2. Power	80,000/yr
3. Water	5,200/yr
4. Maintenance Materials	21,700/yr
5. Polymer*	1,310,000/yr
Total O & M	$1,492,900/yr
Total Annual	$1,731,900/yr
$/ton	$ 40.90/ton

*Includes credit for sludge taken from Primary Sludge Station.

TABLE A-6. COST ESTIMATE SUMMARY FOR PRESSURE FILTRATION DEWATERING OF DIGESTED WASTE ACTIVATED SLUDGE

CONDITIONS	
1. Sludge Quantity	83 tons/day
2. Solids Loading	0.31 lbs/hr-ft^2
3. Sludge Conditioning	240 lbs/ton $FeCL_3$ (9.96 tons/day) 800 lbs/ton CaO (33.2 tons/day) 150 lbs/ton DE (6.5 tons/day)
4. Filtrate Suspended Solids	50 mg/l
5. Total Cake Solids	34% TS
CAPITAL COSTS	
1. 3-7500 ft^2 presses-purchase	$3,350,000
2. Installation & Housing	838,000
3. Contingencies	838,000
4. Contractors Profit	418,800
5. Engineering Fee	670,000
Total Capital	$6,114,800
Annual Capital(7% - 15 yrs)	670,500
OPERATION & MAINTENANCE	
1. Labor	$ 114,000/yr
2. Power	100,000/yr
3. Water	2,000/yr
4. Maintenance Material	61,200/yr
5. Chemical	1,440,000/yr
Total O & M	[illegible],717,200/yr
Total Annual	$2,387,700/yr
$/ton	$ 78.80/ton

TABLE A-7. COST ESTIMATE SUMMARY FOR THERMAL TREATMENT

CONDITIONS

1. Detention Time	30 min
2. Volume	2100 CF
3. Temperature	400°F

CAPITAL COST

Thermal Unit

1. Installed Cost*	$ 7,800,000
2. Housing	600,000
3. Contingencies	1,500,000
4. Engineering Fee	1,200,000
5. Contractors Profit	750,000

Side Stream Treatment+

1. Construction Cost	$ 2,167,000
2. Engineering Fee	433,000
Total Capital	$14,450,000
Annual Capital++	$ 1,521,500

OPERATION & MAINTENANCE

1. Labor	$ 171,000/yr
2. Power***	44,000/yr
3. Fuel	300,000/yr
4. Water	5,000/yr
5. Chemical (HNO_3)	4,000/yr
6. Maintenance Materials	197,000/yr
Total O & M	$ 721,000/yr
Total Annual	$ 2,242,540/yr
$/ton	$ 57.98/ton

* Manufacturers estimate for purchase and installation of boiler, titanium exchangers & reactor decant tank, odor control equipment.

+ Anaerobic Filter 2 day detention time.

# 7% - 15 yrs Thermal Unit, 7% - 25 yrs Anaerobic Filter.

** Includes credit for filter gas.

TABLE A-8. COST ESTIMATE SUMMARY FOR PRESSURE FILTRATION OF THERMAL CONDITIONED WASTE ACTIVATED SLUDGE

CONDITIONS	
1. Sludge Quantity	95 tons/day
2. Solids Loading	0.92 lbs/hr-ft^2
3. Filtrate Suspended Solids	250 mg/l
4. Total Cake Solids	42% TS
CAPITAL COST	
1. 2-4300 ft^2 presses - purchase	$1,100,000
2. Installation & Housing	275,000
3. Contingencies	275,000
4. Contractors Profit	138,000
5. Engineering Fees	220,000
Total Capital	$2,008,000
Annual Capital (7% - 15 yrs)	220,000
OPERATION & MAINTENANCE	
1. Labor	$ 114,000/yr
2. Power	24,000/yr
3. Maintenance Materials	20,000/yr
Total O & M	$ 158,000/yr
Total Annual	$ 378,000/yr
$/ton	$ 10.90/ton

TABLE A-9. COST ESTIMATE SUMMARY FOR VACUUM FILTRATION OF THERMAL CONDITIONED WASTE ACTIVATED SLUDGE

CONDITIONS	
1. Sludge Quantity	95 ton/day
2. Solids Loading	3.0 lbs/hr-ft^2
3. Filtrate Suspended Solids	5000 mg/l
4. Total Cake Solids	35% TS
CAPITAL COSTS	
1. 3-900 ft^2 filters - purchase	$478,000
2. Installation & Housing	119,500
3. Contingencies	119,500
4. Contractors Profit	59,800
5. Engineering Fee	95,600
Total Capital	$872,400
Annual Capital (7% - 15 yrs)	95,700
OPERATION & MAINTENANCE	
1. Labor	$114,000/yr
2. Power	84,000/yr
3. Maintenance Material	10,000/yr
Total O & M	$208,000/yr
Total Annual	$303,700/yr
$/ton	$ 8.80/ton

APPENDIX B

UNIT CONVERSIONS

TABLE B-1. UNIT CONVERSIONS[17]

Customary Unit	Conversion	Metric Unit
BTU/lb	X2.326	kJ/kg
BTU/hr/ft^2	X3.154	J/m^2·s
cfm	X0.4719	l/s
cfm/ft^3	X1.667X10^{-5}	l/m^3·s
^{o}F	(^{o}F-32)/1.8	^{o}C
ft	X0.3048	m
ft^3/lb	X0.06234	m^3/kg
gal	X0.003785	m^3
GPD	X0.003785	m^3/d
gpm	X0.06308	l/s
in	X25.4	mm
in	X0.0254	m
in.Hg	X3.38	kPa
lb	X0.4536	kg
lb/dy	X0.4536	kg/dy
lb/ft^3-dy	X16.02	kg/m^3·d
lb/hr	X0.4536	kg/hr
lb/hr·ft^2	X4.883	kg/m^2·h
lb/ton	X0.500	g/kg
MGD	X0.0438	m^3/s
psi	X6.895	kPa
ton	X907.2	kg
ton/dy	X907.2	kg/d